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Feasibility Assessment of Biogas Generation by Anaerobic Fermentation of Pulp and Paper Wastes

by

ADI Limited

Canada

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ENFOR Project C-300

Feasibility Assessment of Biogas Generation by Anaerobic Fermentation of Pulp and Paper Wastes

by

ADI Limited

Contractor: ADI Limited
1115 Regent Street
Fredericton, N.B.
E3B 4Y2

Scientific Authority:

Ms. W.M. Lyse Morisset
Biotechnology Program Officer
National Research Council of Canada
CISTI Bldg. M-55
Ottawa, Ontario
K1A 0R6

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May 1983

ABSTRACT

This work was completed under a Government of Canada Contract (ENFOR Project C-300) to demonstrate the feasibility of biogas generation by anaerobic fermentation of pulp and paper wastes.

Five bench-scale models were operated for nine months to evaluate the feasibility of anaerobic fermentation of two pulp and paper waste streams. Two downflow filters (DFF), one with cloth and the other with rock media, and an ADI-bulk volume fermenter (BVF) were fed diluted high yield spent sulphite liquor (HYSSL) at average BOD loads of 1.00, 0.81 and 0.42 $kg/m^3.d$, respectively. BOD removals and specific methane productions were 71.0, 67.2 and 73.3% and 0.56, 0.47 and 0.52 m^3 CH_4/kg BOD removed, respectively. Dynamic testing revealed the cloth media DFF biogas production response to shock load and feed cessation.

Newsprint mill primary clarifier sludge (PCS) was fed to two wood media DFF's, one at 35°C and the other at 55°C. COD removals (estimated on the basis of methane production) of 26% and 24%, respectively, were observed. The fibrous nature of the PCS posed difficulties for model operation, gas collection and sample analysis. Improvements in PCS settling and dewatering were noted as a result of anaerobic digestion.

Economic projections for full-scale anaerobic treatment of HYSSL and PCS were made on the basis of model results. The results of these projections indicate that anaerobic fermentation may be an economical means of extracting valuable energy from these waste streams and reducing treatment and disposal costs.

A sulphur balance conducted on the HYSSL cloth DFF model indicated that 66% of the total sulphur in the waste was removed by a crude sulphide scrubber-stripping system. The scrubber-stripper concept along with anaerobic treatment offers very interesting possibilities for chemical and energy recovery and waste treatment.

In view of the encouraging findings in this study, the Canadian Government and the pulp and paper industry are encouraged to sponsor further anaerobic fermentation research at both the bench and pilot level.

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LIST OF ABBREVIATIONS

a	annum
BOD	biochemical oxygen demand
BVF	bulk volume fermenter
C.D.	contact digester
COD	chemical oxygen demand
d	day
DFF	downflow filter
g	gram
GJ	gigajoule
HRT	hydraulic retention time
HYSSL	high yield spent sulphite liquor
kg	kilogram
kJ	kilojoule
kWh	kilowatt hour
l	litre
m	metre
mg/l	milligram per litre
MJ	megajoule
mm	millimetre
PCS	primary clarifier sludge
SRT	solids retention time
SS	suspended solids
SSA	specific surface area
t	metric tonne
TKN	total kjeldahl nitrogen
TS	total solids
TVS	total volatile solids
ug/l	micrograms per litre
VA	volatile acid
VSS	volatile suspended solids

1.0 INTRODUCTION

The energy crisis of the 1970's stimulated a major increase in activity in the field of anaerobic treatment. This was not surprising considering that anaerobic treatment significantly impacts on energy conservation in two ways regarding waste treatment:

(1) the process generates biogas which is a source of energy; and

(2) the process supplements much of the energy which would otherwise be required for aeration and sludge dewatering in aerobic waste treatment.

Inherent in anaerobic fermentation is the production of biogas which is a mixture of methane (CH_4) and carbon dioxide (CO_2). In theory, 0.35 m^3 of CH_4 is generated for every kilogram of chemical oxygen demand (COD) destroyed anaerobically. Typical methane concentrations and gross heating values for biogas are in the range of 60 to 80% and 22 000 to 30 000 kJ/m^3, respectively. The biogas can be collected and utilized to displace fossil fuels.

Of equal importance to many is the large electrical energy savings of anaerobic fermentation when compared with conventional aerobic processes. The electrical energy savings plus the generation of biogas can result in a potential gross energy savings in the order of 23 gigajoules (GJ) per tonne of COD destroyed in the process (worth approximately $95 assuming $5.50/GJ, reference Appendix 1). Thus, assuming typical COD loads of 25-50 t/d removed, anaerobic fermentation systems can be rightfully construed as energy production/conservation systems and not merely an alternate form of waste treatment.

Anaerobic fermentation offers other advantages which can include:

(1) lower capital and operating costs than an equivalent conventional aerobic system;

(2) ability to ferment high-strength, high-solids waste streams;

(3) ability to cope with shock loadings;

(4) low biomass yield plus digestion of waste sludge, hence lower sludge disposal costs;

(5) low nutrient requirements;

(6) a high quality effluent (which further improves as the system "matures") when used as a pretreatment step followed by aerobic polishing (combined removals of 90 to 99% are typical).

Investigations have shown that in many industrial waste treatment applications, anaerobic fermentation can generate 10 to 40% of the fuel requirements for the process plant generating the waste. It is not uncommon for anaerobic systems to generate simple paybacks in the order of 1 to 3 years.

Until recently, anaerobic processes found slow acceptance in industrial applications. This slow acceptance was due to a number of factors, including:

(1) poor performance and past operating problems;

(2) odour problems from anaerobic lagoons;

(3) abundance of cheap energy;

(4) concern over the sensitivity of the organisms to toxic materials and temperature and pH changes;

(5) perceived temperature requirements of 35°C or higher; and

(6) lack of understanding of process variables and control in the case of anaerobic lagoons.

The above factors, questions and concerns have been addressed by the many laboratory, pilot and full-scale anaerobic studies undertaken in the last few years. These studies have also resulted in the development of a number of effective anaerobic fermentation systems (reference Section 5.0). These systems have found a number of applications, especially in the food and drink sector. However, even with all the laboratory and full-scale successes, anaerobic treatment is not, universally, considered tried and proven technology. There is still skepticism, concern and apparent limitations which can only be overcome by more intense research and development programs and, of course, full-scale applications operating successfully over a period of years.

The pulp and paper industry is energy intensive. Most mills are investigating and implementing different schemes for reducing energy costs in order to improve and/or maintain cost competitiveness. These energy conservation schemes have included technologies for the direct combustion, gasification and liquifaction of wood wastes. However, it was only recently that the pulp and paper industry began looking at its liquid waste streams as potential energy sources; normally, where conventional ex-plant treatment is required, these streams are energy sinks.

In the chemical pulping sector, 20 to 50% of the incoming wood solids exit the process in spent cooking liquors. In Canada there are still a number of small pulp mills where liquor incineration and energy recovery is not practised because of its expense; at these mills the spent liquor is a very large source of pollution. However, there is potential to anaerobically convert these spent liquors into major sources of biogas energy. Anaerobic fermentation may, in such cases, prove to be the only cost-effective means of both recovering the energy and dealing with the pollution abatement problem.

The potential of anaerobic treatment in the pulp and paper sector looks promising. The present study, as well as several others by ADI and other organizations, confirm that anaerobic treatment can be successfully applied to a number of problematic waste streams. Furthermore, there are at present two successful full-scale anaerobic systems in existence (1), one operating on bark leachate and the other on waste paper board effluent. There are at least two full-scale systems under construction in Europe for treatment of pulp mill effluents, and two more are in the late planning stages in Canada.

It is essential that Canada continue funding of anaerobic research and development and demonstration programs so that it can remain a world leader in this area of technological development. Otherwise, it will lose its leadership role and may never reap the full benefits of this technology in its industrial sectors.

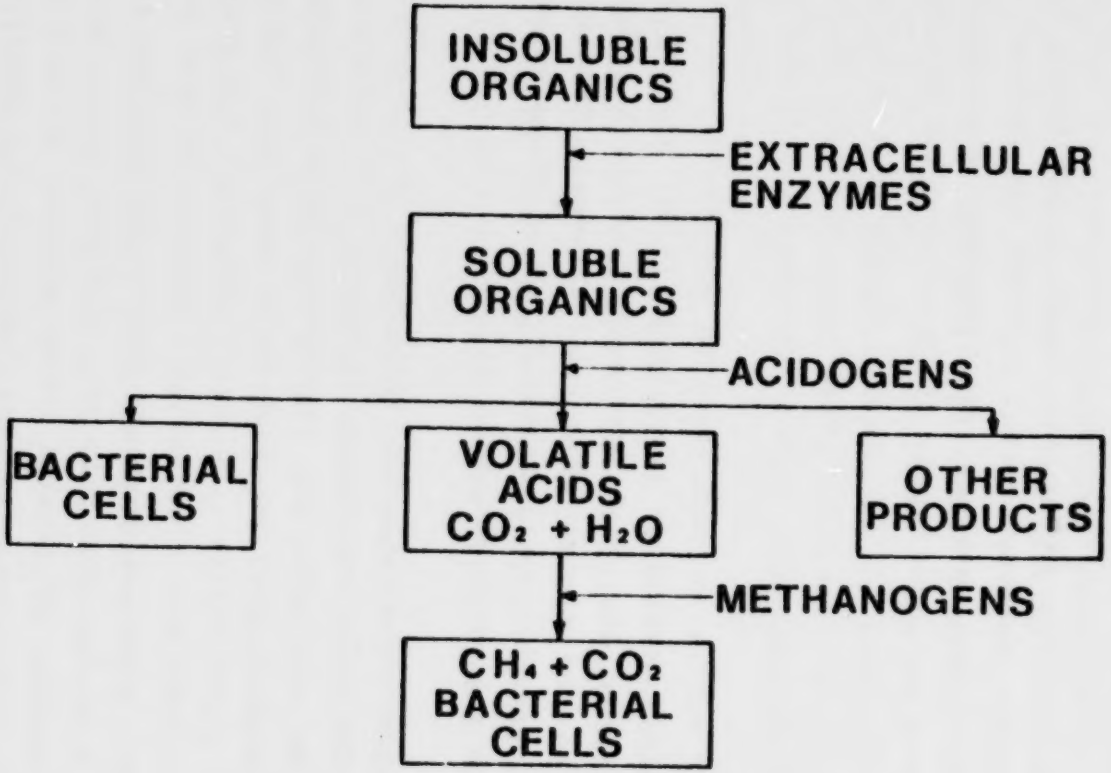
2.0 THEORY AND BACKGROUND INFORMATION

2.1 Theory

Anaerobic fermentation is carried out by a mixed culture of interdependent microorganisms. It is well suited for the digestion of both soluble and insoluble wastes.

In the case of an insoluble substrate, the basic fermentation can be broken down into three main steps, viz., hydrolysis, acid formation and methane formation, which all proceed simultaneously. Enzymatic hydrolysis produces a mixture of sugars, amino acids and long chain fatty acids. These become substrate for the acid-forming group of bacteria which convert them into short chain volatile acids (mainly acetic and propionic). The methane fermentation step is carried out by the methanogens which convert the short-chain volatile acids and other simple compounds to methane and carbon dioxide. It is only during the last stage of the process that waste stabilization actually occurs. The methanogenesis step has very low cell yield and consequently low cell growth rates; it is therefore, normally, the rate limiting step in the fermentation process. The basic mechanism of anaerobic fermentation is illustrated in Figure 2.1.

It is generally considered that any waste which can be stabilized aerobically can also be stabilized anaerobically. This infers that (barring other complications) anaerobic processes could have wide application in the pulp and paper industry.



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Drawing: MECHANISM OF ANAEROBIC TREATMENT		Date: MAR./83	Drawing No.: FIGURE 2.1

ADI 2-187

There are three optimal temperature ranges for growth of anaerobes:

psychrophilic - 5 to 15°C;
mesophilic - 30 to 35°C; and
thermophilic - 50 to 55°C.

Research has shown that the mesophilic range is preferred for most anaerobic systems. However, recent research and full-scale experience indicates that systems can provide satisfactory levels of performance at temperatures of 15 to 20°C and perhaps lower (2, 3).

2.2 Background Information

ADI has been very active in the research and development of anaerobic fermentation technology since early 1978. Much of its early work centered on the food processing industry. These efforts resulted in the design of three full-scale treatment systems using ADI anaerobic technologies. Two of these systems have been built and are performing well.

In 1980 ADI saw the potential for application for anaerobic fermentation to the pulp and paper industry. A literature survey conducted at that time (1) revealed several laboratory and pilot-scale studies and two full-scale anaerobic systems treating pulp and paper wastes. Shortly thereafter ADI undertook a successful anaerobic treatability study on a NSSC pulp mill effluent (4). This was followed shortly by another successful study on a second NSSC pulp mill effluent (5). During this period the results of several other studies on pulp and paper wastes had been completed and published by different researchers (6, 7, 8) and ADI had completed in-house studies on a number of pulp and paper waste streams.

The results of these studies all pointed to the potential for application of anaerobic treatment to many pulp and paper mill effluents and the savings of large amounts of energy (e.g., in the case of one study (5) it was projected from bench-scale results that one particular NSSC mill could save $500 000 per year by displacing natural gas with biogas and another $600 000 per year in electrical (aeration) costs, if it had to provide an equivalent conventional aerobic treatment system.

In late 1981, ADI made application under the ENFOR Program for a bench-scale study. The purpose of the study was to assess the feasibility of biogas generation by anaerobic fermentation of several pulp and paper waste streams. These particular wastes had been found difficult and/or expensive to deal with by conventional means.

On the basis of that submission, the ENFOR Committee awarded ADI a contract, but limited it to the following two waste streams:

(1) high yield spent sulphite liquor (HYSSL); and
(2) primary clarifier sludge (PCS) from a newsprint mill.

The interest in anaerobic fermentation of HYSSL is due to the fact that the conventional sulphite liquor treatment systems are usually difficult to justify economically. This is especially true for many Canadian sulphite mills which are old and not very profitable.

The methods of spent sulphite liquor treatment which have proven most feasible are:

(1) by-products and chemicals recovery;
(2) thermal oxidation with heat and chemical recovery; and
(3) external secondary treatment.

These methods are only feasible when pollution control is a major factor.

The economics for by-products and chemicals recovery and thermal oxidation with heat and chemical recovery are even less favorable in the case of high yield sulphite mills which tend to be small and have lower chemical consumption and less heat recovery potential.

It was felt that anaerobic fermentation of HYSSL may be a cost-effective alternative for liquor treatment. This process has the potential to generate a large amount of biogas from the waste organics, while providing a substantial reduction in the biochemical oxygen demand (BOD) with little expenditure in electrical power.

While extensive work has been done to reduce the burden to industry of sludge disposal, it continues it be a major problem. This is due, in part, to increased industrial production, stricter environmental regulations, and the production of large quantities of waste sludges from secondary waste treatment systems.

Anaerobic fermentation of PCS from newsprint and other pulp and paper making operations is of interest for it can result in a number of important benefits, some of which are:

(1) cellulosic materials are well stabilized by anaerobic treatment and therefore produce an innocuous sludge;

(2) anaerobic treatment can significantly reduce sludge volume through solids digestion and improvement in the dewatering characteristics of sludge;

(3) anaerobic treatment produces valuable biogas.

It was hoped that anaerobic fermentation of PCS would indeed result in all the above benefits and result in a more overall cost-effective sludge disposal technique and strategy for the pulp and paper and other industries.

3.0 OBJECTIVES

As mentioned above the ENFOR Committee, after review of ADI's proposal, recommended study of high yield spent sulphite liquor (HYSSL) and primary clarifier sludge (PCS) from a newsprint mill. The purpose of the study was:

(1) in the case of the HYSSL, to assess the feasibility of anaerobic fermentation for production of biogas and concomitant reduction of pollution loading; and

(2) with PCS, to assess the potential for biogas production, reduction in sludge volume and the improvement in the dewatering characteristics of the sludge.

From the results of the study and for hypothetical full-scale mill situations, preliminary capital and annual cost and projected savings estimates were to be developed for both waste streams. This information would allow a simple payback to be calculated for each waste. This information would be used to assess the merits of further study.

The technical feasibility for anaerobic fermentation of each waste was assessed through a nine month bench-scale study. The following bench-scale models and technologies were used in this study:

(1) high yield spent sulphite liquor - all operated at 35-37°C.
   a. downflow filter (NRC technology), rock media
   b. downflow filter (NRC technology), cloth media
   c. bulk volume fermenter (ADI technology)

(2) primary clarifier sludge
   a. downflow filter (NRC technology), wood media - operated at 35-37°C
   b. downflow filter (NRC technology) wood media - operated at 55-57°C.

4.0 WASTE CHARACTERISTICS

4.1 Spent Sulphite Liquor

The spent sulphite liquor used for the study was supplied by a high-yield, sodium base sulphite mill. The waste was shipped to ADI's lab in 100-200 l batches and subsequently stored in 15 l plastic containers and frozen to minimize possible degradation and changes. Waste was thawed, as required, for model operation. HYSSL was diluted with water to prepare model feeds (details, Section 7.0). The results presented in Table 4.1 represent the ranges and average values from a series of three complete analyses of three different shipments of raw spent liquor.

4.2 Primary Clarifier Sludge

The primary clarifier sludge used for the study was supplied by a newsprint mill in Eastern Canada. The PCS was shipped regularly to ADI's lab in 50-80 l batches; it was strained with a 9.5 mm screen to remove large solids that would cause operating problems in the bench models and stored at room temperature. Ideally the sludge should have been stored at 4°C, but this was not possible. It was concluded that short-term room temperature storage would not be a serious problem since the PCS was largely, difficult-to-degrade wood fibre. Table 4.2 presents the ranges and average values from two, complete analyses done on separate shipments of raw PCS.

TABLE 4.1 SPENT SULPHITE LIQUOR CHARACTERISTICS[1,2]

Parameter	Range	Average[3]
BOD	9 020 - 15 100	12 400
COD	75 000 - 93 000	82 300
SS	340 - 1 480	910
VSS	260 - 1 140	700
TKN	25.3 - 78.9	48.1
NH_3-N	LT 1.0	LT[4] 1.0
Total P	4.0 - 6.5	4.8
Total S[5]	5 000 - 8 100	6 200
SO_3	4 090	4 090[6]
SO_4	2 200 - 2 870	2 480
Ca	134 - 150	140
Cd	LT 0.1	LT 0.1
Cr	0.2	0.2
Cu	0.20 - 0.28	0.24
Fe	2.6 - 4.0	3.3
Hg	LT 0.2	LT 0.2
K	68 - 101	88
Mg	21 - 27	23
Na	5 240 - 6 960	6 170
Ni	0.3 - 0.5	0.4
Pb	0.8	0.8
Zn	1.7 - 2.5	2.0

Notes:

1. On as-received (raw) samples, i.e., with no yeast extract, nutrients, etc. added.
2. All results expressed as mg/l, except Hg, as ug/l.
3. Averages of 3 sets of complete analyses.
4. LT = less than.
5. Details Section 8.5
6. Result based on single sample analysis.

TABLE 4.2 PRIMARY CLARIFIER SLUDGE CHARACTERISTICS[1,2]

Parameter	Range	Average[3]
COD	12 000 - 14 000	13 000
TS	10 400 - 10 800	10 600
TVS	10 000 - 10 400	10 200
TKN	12.7 - 15.6	14.2
NH_3-N	LT[4] 1	LT 1
Total P	LT 1 -2	1
SO_4	300 - 420	360
Ca	20.5 - 25.8	23.2
Cd	LT 0.01	LT 0.01
Cr	LT 0.02	LT 0.02
Cu	0.04 - 0.38	0.21
Fe	1.4 - 2.7	2.1
Hg	LT 0.2	LT 0.2
K	3.6 - 5.1	4.4
Mg	2.4 - 3.0	2.7
Na	91 - 140	116
Ni	LT 0.1	LT 0.1
Pb	LT 0.05	LT 0.05
Zn	1.1 - 3.1	2.1

Notes:

1. On as-received (raw) samples, i.e., with no yeast extract, nutrients, etc. added.
2. All results expressed as mg/l, except Hg, as ug/l.
3. Averages of 2 sets of complete analyses.
4. LT = less than.

5.0 ANAEROBIC TECHNOLOGIES

A number of advanced (or high-rate) anaerobic processes have been developed specifically for the treatment of industrial wastewaters. Several of these are now available commercially. These systems differ from the conventional anaerobic (municipal type) digesters in that they attempt to maximize the solids retention time (SRT). This can be accomplished by sludge recycle, in the case of suspended growth systems, or providing media for biomass attachment, in fixed film systems. By providing a high SRT (i.e., high concentrations of microorganisms), high volumetric loading rates and performance can be achieved without micro-organism washout - even with very low hydraulic retention times (HRT).

The first anaerobic system to be applied to industrial wastewaters was the anaerobic lagoon. The anaerobic lagoon is not considered an advanced technology because of its low loading rate. It has not been as successful as it could have been for reasons listed below:

(1) poor design;
(2) unfavourable operating conditions; and
(3) improper startup.

However, an adaptation of the lagoon, the ADI-Bulk Volume Fermenter (ADI-BVF) has recently been used very successfully in two full-scale applications with food processing waste (3,9). The ADI-BVF differs from the anaerobic lagoon for the following reasons:

(1) it utilizes a floating membrane cover for biogas collection and temperature and odour control;

(2) it incorporates biogas utilization;
(3) it employs effluent and sludge recycle to enhance performance;
(4) it provides on stream sludge removal;
(5) it utilizes special inlet and outlet arrangements;
(6) it has provision for alkalinity and pH control; and
(7) includes a monitoring system.

Anaerobic filters are considered high-rate or advanced systems. Operational modes have included downflow (10), upflow (11,12) and horizontal flow (13). They use media to provide surface area and interstitial space to achieve high concentrations of biomass within the reactor. Examples of media used in bench and full-scale filters include, rock, clay tiles, plastics, glass, fabric and wood. The specific surface areas (SSA) and porosity utilized to date have ranged from approximately 50 to 200 m^2/m^3 and 40 to 97 percent, respectively. In full-scale application the tendency has been to employ media with SSA of 30 to 100 m^2/m^3 as a compromise between performance and ease of media cleaning.

Advanced systems, other than the anaerobic filter (some of which have only been used on a laboratory scale) now include, but are not limited to:

(1) the anaerobic contact process (7);
(2) the upflow anaerobic sludge blanket (14);
(3) the two-stage reactor (15,16);
(4) NRC downflow filter (10);
(5) anaerobic attached film expanded bed (17);
(6) anaerobic fluidized bed (18); and
(7) the ADI-HAF (horizontal-mode anaerobic filter) (13).

The development of anaerobic fermentation systems is proceeding along two lines, viz., the high-rate and low-rate routes (with

emphasis on the former). Although, strictly speaking, no firm definitions of high and low-rate systems exist; high-rate systems may be regarded as those designed to be loaded at 5 kg COD/m^3.d, or higher (depending upon the waste to be treated).

The high-rate systems were developed in an effort to reduce reactor volume and thereby reduce capital costs. This reduction in cost is partially offset by an increase in system complexity and need for a higher degree of control. The experience with such systems has been quite favorable and a number of full-scale systems have been installed.

Low-rate systems have also evolved from an attempt at keeping the capital cost as low as possible. However, the designs have also endeavoured to keep the system as simple and robust as possible, consistent with a high degree of system performance. Usually low-rate systems are characterized by inexpensive earthen construction, floating membrane covers for biogass collection, temperature and odor control and a minimum of mechanical components. Low-rate systems can usually be designed to provide the same level of performance as high-rate systems. In addition, they have the advantage of being able to handle high solids loading while providing a high degree of digestion of these solids. This digestion function is accomplished by the extremely high solids retention time in this type of reactor. This ability to handle and digest large amounts of solids has been used to advantage in at least two low-rate full-scale designs (3,19). In both these cases, the low-rate reactor doubled as both primary and secondary treatment systems. This eliminated the need for a primary clarifier and associated sludge dewatering and disposal, while providing a high degree of solids stabilization. The major disadvantage of the low-rate system is the large land requirement.

Canada has been one of the world leaders in the development of anaerobic fermentation technology. Much of this distinction is the result of the work at the NRCC laboratories. This laboratory has been responsible for the development of the NRC downflow filter (DFF) technology. Because of its downflow configuration the DFF is well-suited to handle wastes with high concentrations of suspended solids. In addition, this type of filter has demonstrated its effectiveness on a number of feed streams and over a wide range of operating conditions.

For the present study both the NRC downflow filter and ADI-BVF technologies were used so a comparison could be made.

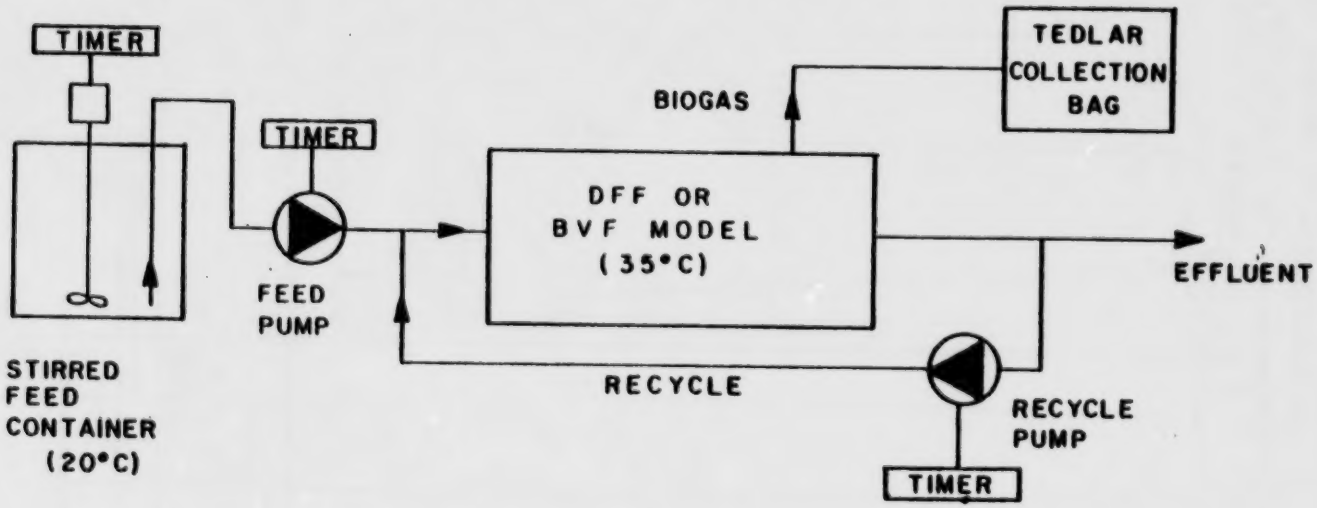
6.0 BENCH-SCALE MODELS

6.1 Spent Sulphite Liquor Downflow Filters

Two of the three models used to study HYSSL treatabiity were DFF's. Figure 6.1 illustrates schematically the operation of these models. While the models were operated at 35°C (in a temperature-controlled room), the feed container was kept at 20°C to reduce waste degradation in the container. The feed mixer and the peristaltic feed and recycle pumps were all actuated by a single timer. New feed batches were prepared weekly.

The two DFF models were constructed of 102 mm inside diameter (i.d.) plexiglas cylinders (each with an overall height of approximately 900 mm) as illustrated in Figure 6.2. The rock media used in one of the DFF's consisted of hand graded 25-40 mm diameter rounded stones having a porosity of approximately 0.42. The rocks were supported by a perforated plate in the model. The top surface of the rocks was approximately 45 mm from the top (inside) of the model. The total (total = rock + liquid) volume of the model was approximately 5.9 l and the liquid volume was approximately 2.6 l. It was estimated that the rock media provided a specific surface area (SSA) of approximately 90 m^2/m^3 based on the total volume of the model (5.9 l).

The second DFF utilized a geotextile fabric (needle punched polyester, approximately 3 mm thick, supplied by Texel Ltd.) as the biomedia. The fabric was stretched on a framework of aluminum rods and nylon rings as illustrated in Figure 6.3. The total volume (total = cloth + support + liquid) of the model was 5.9 l and the liquid volume was 5.4 l. The surface area of the cloth was approximately 0.68 m^2 giving a SSA of 115 m^2/m^3 (based on the total volume of 5.9 l). The cloth and rock media reactors were identical, apart from the media used.



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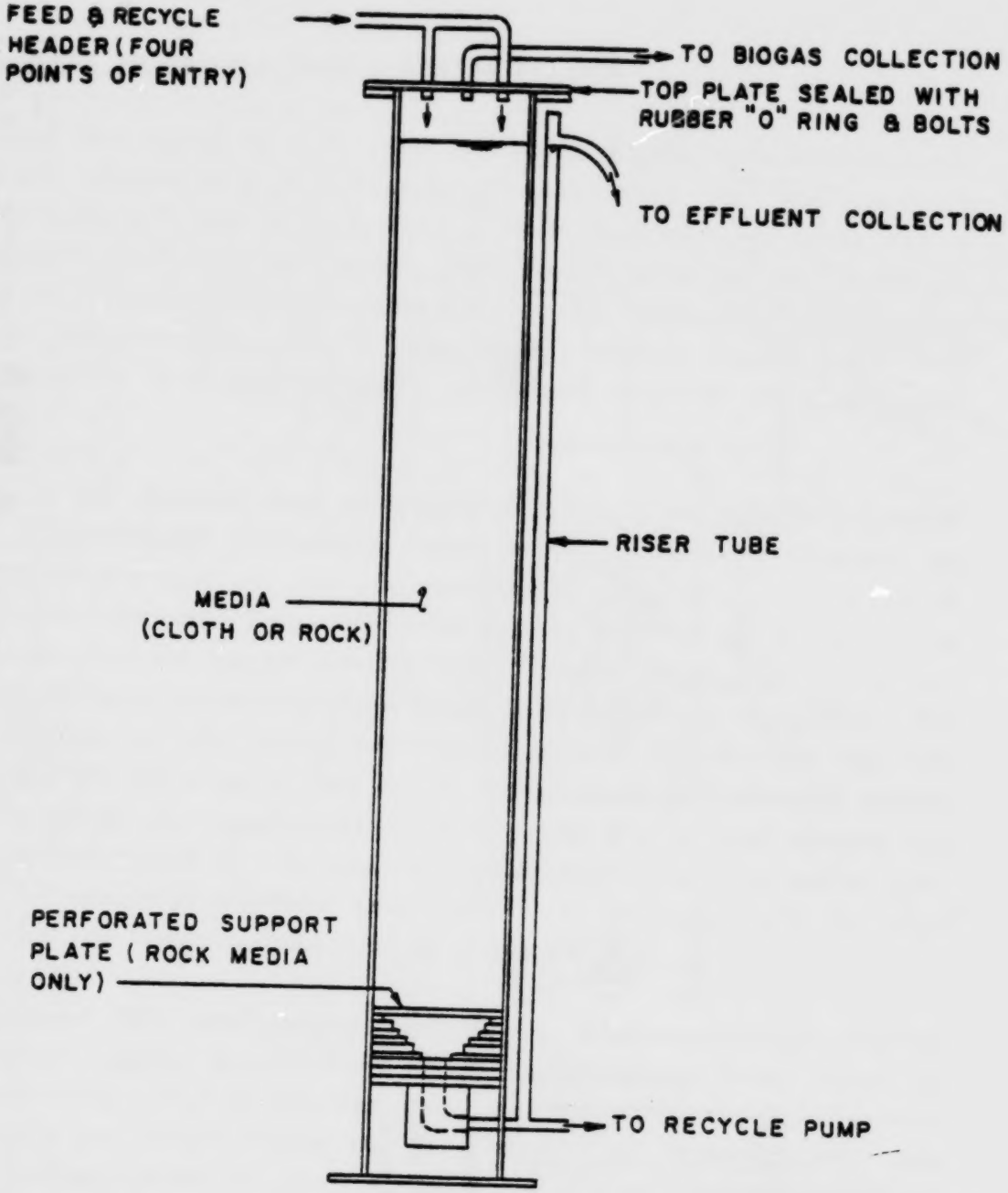
Drawing: SCHEMATIC OF HYSSL MODELS		Drawing No.: FIGURE 6.1		

ADI 2-187


FEED & RECYCLE HEADER (FOUR POINTS OF ENTRY)
TO BIOGAS COLLECTION
TOP PLATE SEALED WITH RUBBER "O" RING & BOLTS
TO EFFLUENT COLLECTION
RISER TUBE
MEDIA (CLOTH OR ROCK)
PERFORATED SUPPORT PLATE (ROCK MEDIA ONLY)
TO RECYCLE PUMP

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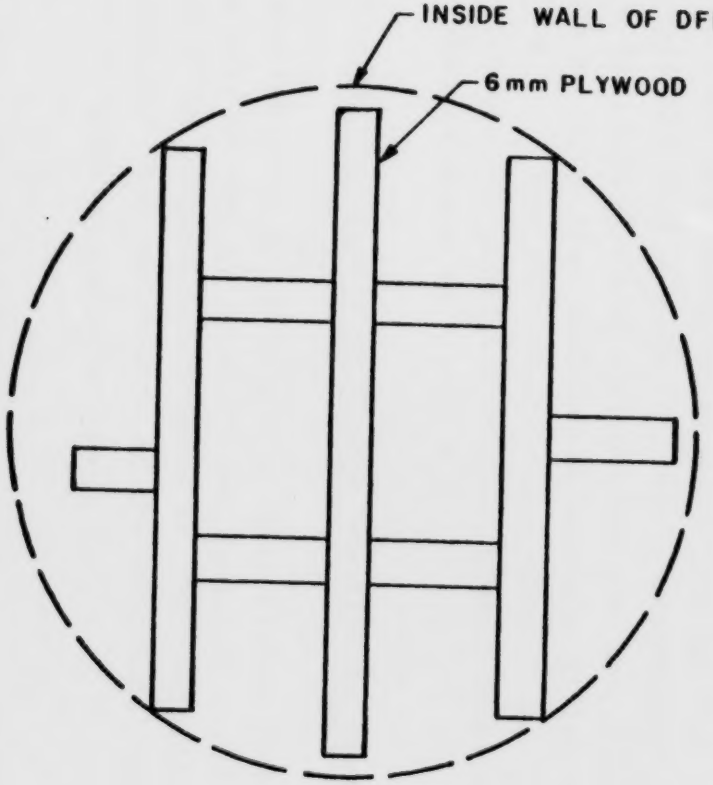
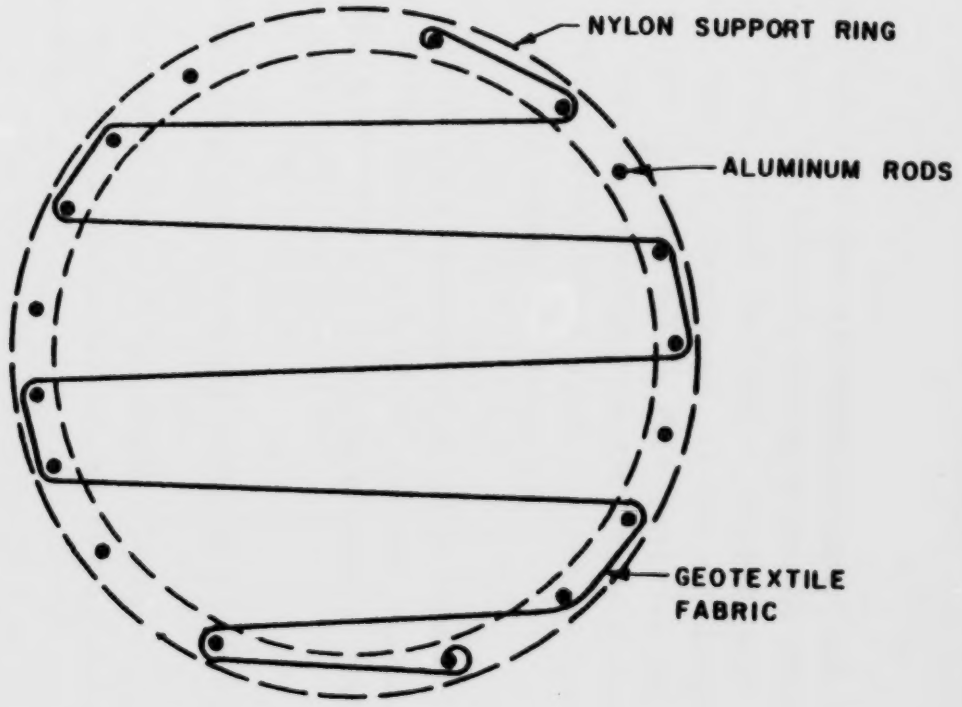
Drawing: SPENT LIQUOR DOWNFLOW FILTER - ELEVATION		Drawing No.: FIGURE 6.2		



WOOD MEDIA - PCS DFF's



CLOTH MEDIA - HYSSL DFF

Job
ENFOR - ANAEROBIC FERMENTATION OF PULP & PAPER WASTES

Drawing
MEDIA CROSS SECTIONS
HYSSL CLOTH MEDIA & WOOD MEDIA

ADI Limited
Engineers, Architects
Management Consultants
Planners, Surveyors
(506) 454 4481
1115 Regent Street,
Fredericton, N.B.

Drn. by	Scale	Date
B.D.L.	1:1	MAR./83

Drawing No.
FIGURE 6.3

ADI 2-187

On day 219 a biogas scrubber-stripper was added to the cloth media DFF. The scrubber-stripper system consisted of a gas pump and 2 litre flask (scrubber) filled with caustic solution, through which a continuous stream of biogas was bubbled for removal of CO_2 and H_2S. This purified gas was then bubbled (coarse bubbles) through the DFF (from bottom to top) to strip H_2S. This system would serve to reduce any stress in the DFF caused by high levels of sulphide.

Model covers were equipped with "O" rings and bolted to a mating flange to produce a gas-tight system. Liquid level within the DFF models was controlled by an external riser tube, with a vent to prevent siphoning.

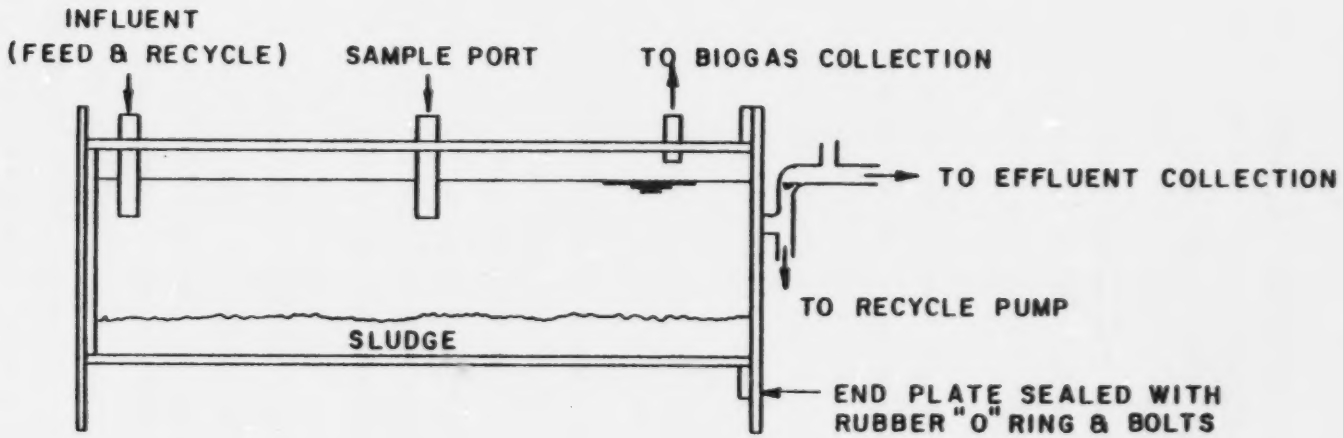
6.2 Spent Sulphite Liquor Bulk Volume Fermenter

The third model used in the HYSSL treatability study was an ADI-BVF. The normal operation of the BVF is shown schematically in Figure 6.1.

The BVF was constructed of a 152 mm i.d. plexiglas cylinder, approximately 500 mm in overall length as illustrated in Figure 6.4. Liquid level in the BVF was maintained at approximately 25 mm from the top inside surface giving a liquid volume of 7.8 l. The end covers were made liquid and gas-tight with suitable "O" rings and bolts. Liquid level in the BVF was regulated as shown in Figure 6.4.

6.3 Primary Clarifier Sludge Downflow Filters

Treatability of primary clarifier sludge (PCS) was studied using a pair of DFF's. One DFF was operated at 35°C and the other at 55°C to compare performance in the mesophilic and thermophilic ranges. The temperature of the 55°C model was maintained by use of an external heating tape.



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Drawing: SPENT LIQUOR - ADI-BVF ELEVATION		Drawing No.: FIGURE 6.4		

ADI 2-109

The DFF's were virtually identical to those used for the HYSSL study, as illustrated in Figure 6.5. The main difference was in the type of media used, in this case, wood.

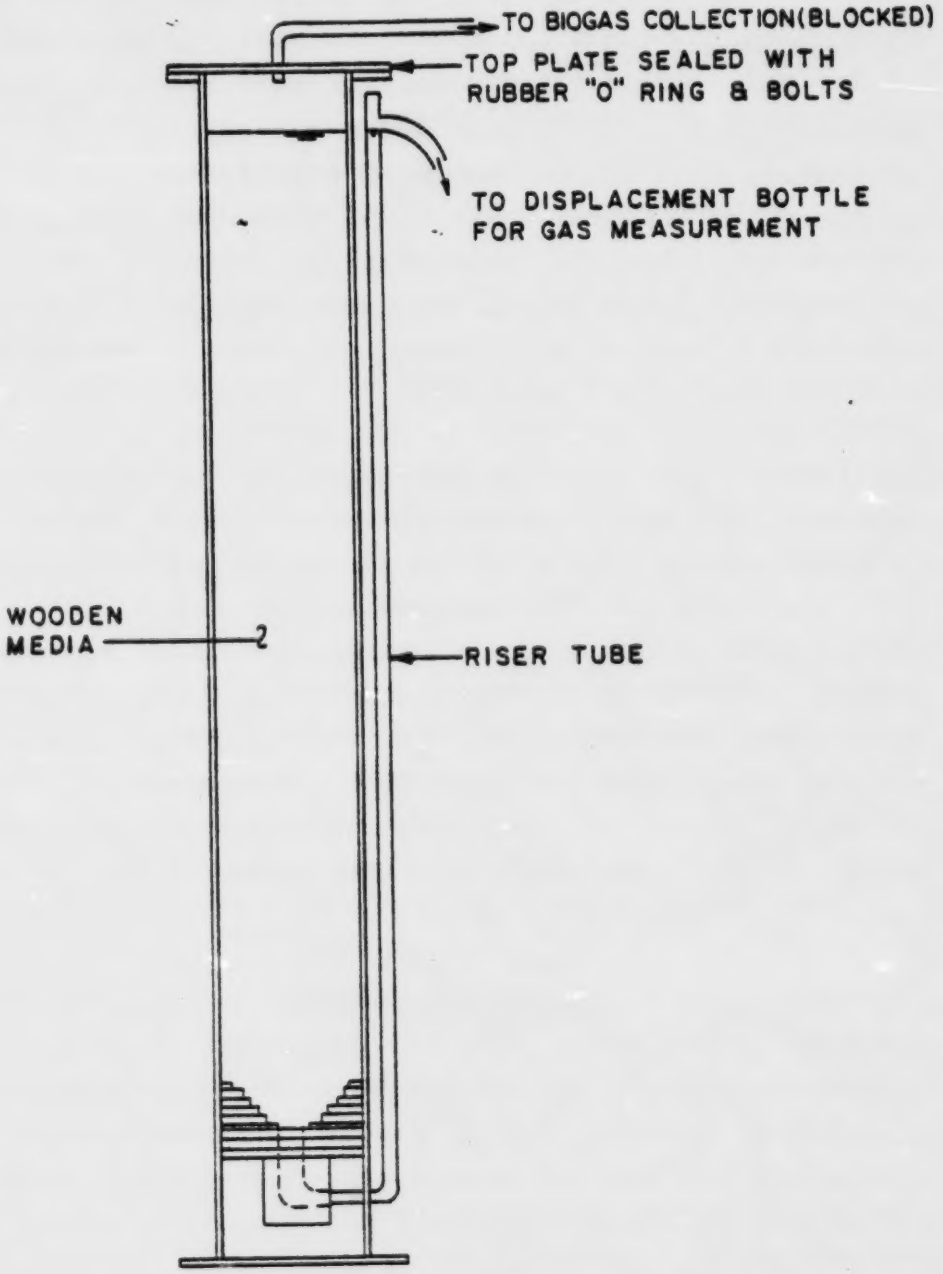
A cross-section of the PCS DFF wood media is shown in Figure 6.3. This wood media supplied a surface area of 0.49 m^2 giving a SSA of 84 m^2/m^3 (based on the total volume).

Initially the media configuration allowed smaller holes (and a higher SSA), but this arrangement was abandoned, due to clogging problems. The total (liquid + wood) and liquid volumes were 5.8 l and 4.2 l, respectively. The reactors were batch-fed daily.

6.4 Biogas Collection

Biogas from all five of the anaerobic models was collected, measured and analyzed to monitor system performance. For the HYSSL models the biogas was collected in Tedlar (polyvinyl fluoride) gas sampling bags (8 l capacity) prior to measurement. Gas measurement was by displacement of an acidified salt solution, as specified in Standard Methods (20).

Initial attempts to measure biogas production in the PCS DFF's were foiled by the fibrous nature of the sludge. However, after a number of modifications a fairly accurate measurement was finally accomplished using displacement of model contents. After each daily feeding the model was sealed and as biogas was produced it forced liquid (digested PCS) from the model into a measuring vessel. The volume displaced was recorded as the daily gas production and the displaced liquid was returned to the model. While it was recognized that this procedure had the shortcoming of not being able to collect gas from the displaced reactor contents, it was sufficiently accurate for the purpose. Biogas was collected for analysis in Tedlar gas sampling bags.



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Drawing: PRIMARY CLARIFIER SLUDGE DOWNFLOW FILTER - ELEVATION		Drawing No.: FIGURE 6.5		

7.0 MODEL OPERATIONS

7.1 Seeding and Startup

Each of the four DFF's used in this study was seeded with 0.5 l of anaerobic sludge. The seed sludge was a blend (in equal portions) of sludges from three previous studies (two food-processing wastes and one pulp and paper waste). The seed sludge had a total solids (TS) and total volatile solids (TVS) of 5.9% and 2.4%, respectively. The ADI-BVF was seeded with 1.5 L of the same seed sludge.

The initial feed for all models was a synthetic waste mixture made with vinegar. The mixture consisted of vinegar (initially to give COD of 1000 mg/l) yeast extract (0.5 g/l), $NaHCO_3$ (1 g/l) and added N (as NH_4CL) and P (as KH_2PO_4) throughout the study to give a COD:N:P ratio of 500:5:1 (assuming no background nutrients in the waste or from the yeast extract). During startup the DFF's were all fed 0.7 l/d (with 1.4 l/d recycle) and the BVF fed 0.2 l/d (with 0.4 l/d recycle). The COD of the vinegar solution was gradually increased during startup and on day 42 of operation HYSSL was introduced to the feed (reference Tables 7.1, 7.2, 7.3).

The feed concentration for the HYSSL models was increased so that by day 70 they were fed 75% HYSSL (25% H_2O). Shortly thereafter (day 73) the models were shut down due to high volatile acids ($\geq$ 2000 mg/l) levels and the feed subsequently diluted (to 6%) and the feed strength gradually increased again. Tables 7.1, 7.2 and 7.3 detail the operation of the BVF, rock-media DFF and cloth-media DFF, respectively.

Feed concentration to the PCS DFF's reached 100% (at 0.7 1/d flow) on day 77, and was maintained for the remainder of the study. The feed rate was reduced from 0.7 1/d to 0.3 1/d on day 182 to increase retention time and improve digestion. Table 7.4 presents the operational information for DFF No. 1 (55°C) and DFF No. 2 (35°C). During the entire startup period all models were operated at 35°C (except PCS DFF No. 1, at 55°C).

7.2 Steady-State Operations

Following the startup period, all models were operated at more or less "steady-state" (day 182 onwards for PCS DFF's and HYSSL BVF and day 147 onwards for HYSSL DFF's), with variations (as detailed in Tables 7.1 - 7.4) according to individual model performance and needs. During steady-state operations all models were operated at 35°C (except PCS DFF No. 1, at 55°C). The basic "design" BOD loads (struck before the start of the study) for the HYSSL models were 2.0 $kg/m^3.d$ for the DFF's and 0.5 $kg/m^3.d$ for the BVF; further, it was assumed that full strength HYSSL could be used. During the study it was found that the waste had to be diluted (and flows increased to maintain loading) to reduce system stress to acceptable limits. The recycle rate remained at 2 to 1 throughout the study.

Initially, the objective with the PCS DFF'S was to achieve a COD/ TVS loading of approximately 1.3/1.0 $kg/m^3.d$. This objective was never fully realized because of the many operating problems encountered.

The mode of HYSSL model operation is presented schematically in Figure 6.1. The PCS models had to be batch-fed, daily, due to the fibrous nature of the PCS (caused clogging in pump lines). In fact, on day 150 the PCS wood media was also modified (pore spaces enlarged) to reduce clogging within the model.

TABLE 7.1 BVF OPERATION

Day	Event/Comments
1	BVF seeded, flow set at 0.2 l/d (with 2:1 recycle). Initial feed - vinegar solution (synthetic waste), with COD ~ 1000 mg/l, N and P added to give COD:N:P = 500:5:1 plus 0.5 g/l yeast extract, 0.5 g/l $Ca(OH)_2$ and 1 g/l $NaHCO_3$.
7	Feed COD increased to 1500 mg/l
14	Feed COD increased to 2000 mg/l
21	Feed COD increased to 3000 mg/l
28	Feed COD increased to 4000 mg/l
42	12% HSSL (by volume) added to vinegar feed
50	Feed increased to 18% HYSSL (plus vinegar feed), $NaHCO_3$ used to neutralize to pH 7(~2 g/l)
56	Feed increased to 25% HYSSL (plus 75% water-vinegar feed deleted from day 56 to end of study)
63	Feed increased to 50% HYSSL
70	Feed increased to 75% HYSSL
73	Shutdown due to high VA
75	Flow resumes at 0.2 l/d
78	Feed diluted to 6% HYSSL and flow rate increased to 0.9 l/d to "rinse" system
85	Feed increased to 25% HYSSL at 0.2 l/d flow
91	Flow (only) increased to 0.4 l/d
93	From day 93 to 129 no total sulphide tests done - chemicals out of stock
99	Therapeutic dose of 15 ml/d of seed sludge added for next 14 days
105	Feed increased to 30% HYSSL at 0.4 l/d
112	Feed (only) increased to 35% HYSSL
122	Feed concentration decreased to 23% HYSSL, flow increased to 0.6 l/d (in order to reduce effluent VA and H_2S, while keeping loading constant)

TABLE 7.1 BVF OPERATION (Continued)

Day	Event/Comments
140	Feed concentration decreased to 20% HYSSL, flow increased to 0.8 1/d, no yeast additions (out of stock) and N added as NH_4HCO_3
147	Flow (only) increased to 0.9 1/d
154	Flow (only) increased to 1.0 1/d, yeast (0.5 g/1) reintroduced
164	Test procedure for total sulphide changed from HACH to Standard Methods.
182*	Feed concentration increased to 25% HYSSL, flow decreased to 0.8 1/d and $NaHCO_3$ again used (2 g/1)
191	From day 191 to 206 no tests done for total sulphide, pH, alkalinity or VA - (holidays)
204	Feed concentration increased to 30% HYSSL, flow at 0.8 1/d, N & P still added to give (at least) COD:N:P = 500:5:1
210	Feed concentration increased to 35% HYSSL, flow decreased to 0.7 1/d and $NaHCO_3$ to 1.5 g/1
231	Started using new shipment of HYSSL for feeds, note high total sulphide, BOD and COD. Feed concentration and flow unchanged (held constant until end of study) and $NaHCO_3$ decreased to 1.25 g/1
271	Data collection complete.

Notes: 1. Q = 0.2 1/d from day 1 to day 78
2. vinegar feed deleted from day 56 to end of study.

*day 182 to end called "steady-state" period for BVF.

TABLE 7.2 ROCK MEDIA DFF OPERATION

Day	Event/Comments
1	DFF seeded, flow set at 0.7 l/d (with 2:1 recycle). Initial feed - vinegar solution (synthetic waste), with COD $\sim$ 1000 mg/l, N and P added to give COD:N:P = 500:5:1 plus 0.5 g/l yeast extract, 0.5 g/l $Ca(OH)_2$ and 1 g/l $NaHCO_3$.
7	Feed COD increased to 1500 mg/l
13	Lost approximately one-third of model liquid contents due to broken recycle line; level restored with water
14	Feed COD increased to 2000 mg/l
21	Feed COD increased to 3000 mg/l
28	Feed COD increased to 4000 mg/l
42	12% HYSSL (by volume) added to vinegar feed
50	Feed increased to 18% HYSSL (plus vinegar feed), $NaHCO_3$ used to neutralize to pH 7(2 g/l)
56	Feed increased to 25% HYSSL (75% water, vinegar feed deleted from day 56 to end of study)
63	Feed increased to 50% HYSSL
70	Feed increased to 75% HYSSL
73	Shutdown due to high VA
75	Flow resumes at 0.7 l/d
78	Feed diluted to 6% HYSSL (high VA) and flow increased to 2.8 l/d to "rinse" system
85	Feed increased to 25% HYSSL at 0.7 l/d flow
91	Flow (only) increased to 1.4 l/d
93	Model shutdown due to high VA. From day 93 to 129 no total sulphide tests done - chemicals out of stock.
96	Flow resumed at 1.4 l/d
98	Feed concentration decreased to 6% HYSSL, flow to 0.7 l/d
99	Model was "rinsed" with 10 l of 6% HYSSL. Therapeutic dose of 15 ml/d of seed sludge added for next 14 days.

TABLE 7.2 ROCK MEDIA DFF OPERATION (Continued)

Day	Event/Comments
105	Feed concentration increased to 12% HYSSL at 0.7 1/d. Increased $NaHCO_3$ to 4 g/l
112	Feed concentration increased to 18% HYSSL
120	Feed concentration increased to 22% HYSSL
126	Feed concentration increased to 30% HYSSL, at 0.7 1/d
140	Feed concentration increased to 35% at 0.7 1/d, no yeast addition (out of stock) and N added as NH_4HCO_3
147	Flow (only) increased to 0.9 1/d - "steady-state" period begins
154	Flow (only) increased to 1.0 1/d, yeast (0.5 g/l) reintroduced
164	Test procedure for total sulphide changed from HACH to Standard Methods.
167	Through accidental siphon the model lost nearly all of its liquid contents. Model was refilled with effluent and water.
168	Feed concentration decreased to 20% HYSSL at 1.0 1/d
175	Flow increased to 1.1 1/d
182	Feed concentration increased to 35% HYSSL, at 0.8 1/d and $NaHCO_3$ reintroduced (2 g/l)
189	Flow increased to 0.9 1/d
191	From day 191 to 206 no tests done for total sulphide, pH, alkalinity or VA - (holidays)
204	Feed concentration increased to 40% HSSL at 0.9 1/d
210	Flow increased to 1.0 1/d and $NaHCO_3$ decreased to 1.75 g/l
224	Flow increased to 1.1 1/d and $NaHCO_3$ dropped to 1.5 g/l
231	Started using new shipment of HYSSL for feeds, note high total sulphide, BOD and COD. Flow increased to 1.2 1/d and $NaHCO_3$ decreased to 1.25 g/l

TABLE 7.2 ROCK MEDIA DFF OPERATION (Continued)

Day	Event/Comments
233	Model shutdown due to high VA
236	Model restarted
240	Model shutdown due to high VA
241	Fed 6 l of buffered (3 g/l $NaHCO_3$) water to model
245	Feed resumed (40% HYSSL at 1.2 l/d)
248	Feed off due to high VA
250	Feed resumed at 1.2 l/d
252	Fed 6 l of buffered water to model and diluted feed to 20% HYSSL at 1.2 l/d
255	Flow decreased to 0.6 l/d
258	Model shutdown (permanently) due to high VA

TABLE 7.3 CLOTH MEDIA DFF OPERATION

Day	Event/Comments
1	DFF seeded, flow set at 0.7 l/d (with 2:1 recycle). Initial feed - same as BVF model.
7	Feed COD increased to 1500 mg/l
14	Feed COD increased to 2000 mg/l
21	Feed COD increased to 3000 mg/l
28	Feed COD increased to 4000 mg/l
42	12% HYSSL (by volume) added to vinegar feed
50	Feed increased to 18% HYSSL (plus vinegar feed), $NaHCO_3$ addition 2 g/l
56	Feed increased to 25% HYSSL (75% water, vinegar deleted from feed starting day 56)
63	Feed increased to 50% HYSSL
70	Feed increased to 75% HYSSL
73	Model shutdown due to high VA
75	Feed resumed at 0.7 l/d
76	Model shutdown due to high VA
78	Feed resumed at 6% HYSSL (high VA) and 2.8 l/d (to "rinse" model)
85	Feed increased to 25% HYSSL at 0.7 l/d
91	Flow increased to 1.4 l/d
93	Model shutdown due to high VA. From day 93 to 129 no total sulphide tests done - chemicals out of stock.
96	Feed resumed at 1.4 l/d
97	Model shutdown
98	Feed decreased to 6% HYSSL at 0.7 l/d
105	Feed concentration increased to 12% HYSSL at 0.7 l/d. Increased $NaHCO_3$ to 4 g/l
112	Feed concentration increased to 18% HYSSL at 0.7 l/d.

TABLE 7.3 CLOTH MEDIA DFF OPERATION (Continued)

Day	Event/Comments
120	Feed concentration increased to 22% HYSSL.
126	Feed concentration increased to 30% HYSSL
140	Feed concentration increased to 35% at 0.7 l/d. No yeast addition and N added as NH_4HCO_3
147	Flow (only) increased to 0.9 l/d, "steady-state" period begins
154	Flow (only) increased to 1.0 l/d, yeast reintroduced (0.5 g/l)
164	Test procedure for total sulphide changed from HACH to Standard methods.
168	Flow (only) increased to 1.1 l/d
175	Flow (only) increased to 1.2 l/d
182	$NaHCO_3$ decreased from 4 to 2 g/l
191	From day 191 to 206 no tests done for total sulphide, pH, alkalinity or VA (holidays)
204	Feed concentration increased to 40% HYSSL at 1.2 l/d
210	$NaHCO_3$ decreased to 1.75 g/l
212	Model shutdown due to high VA. Data collection disrupted - see Figures 8.7 - 8.11.
217	Model contents diluted by 50% with water
219	Gas scrubber connected to model to reduce H_2S in gas. Scrubber gas pumped through model to reduce H_2S in reactor.
220	Feed resumed with 40% HYSSL at 1.2 l/d
224	$NaHCO_3$ decreased to 1.5 g/l
231	Started using new shipment of HYSSL for feeds, note total sulphide, BOD and COD. $NaHCO_3$ decreased to 1.25 g/l

TABLE 7.3 CLOTH MEDIA DFF OPERATION (Continued)

Day	Event/Comments
261	Feed off, dynamic testing
264	Feed resumed
268	Feed off, dynamic testing
271	Feed resumed
274	Shock load applied for 24 h, 40% HYSSL at 3.6 l/d
275	Feed returned to 1.2 l/d
279	Model shut down (high VA)

TABLE 7.4 PCS DFF NO. 1 & NO. 2 OPERATION

Day	Event/Comments
1	DFF seeded, flow rate set at 0.7 l/d (with 2:1 recycle). Initial feed - same as BVF and HYSSL DFF's
7	Feed COD increased to 1500 mg/l
14	Feed COD increased to 2000 mg/l
21	Feed COD increased to 3000 mg/l
28	Feed COD increased to 4000 mg/l
42	12% HYSSL (by volume) added to vinegar feed (HYSSL used as no PCS available)
50	Feed increased to 18% PCS (plus vinegar feed)
56	Feed increased to 25% PCS (75% water, vinegar deleted from feed, day 56 onward)
63	Feed increased to 50% PCS
64	Feed diluted to 25% PCS and flow increased to 1.4 l/d, to reduce clogging
70	Feed increased to 2.1 l/d
76	Pumps disconnected from models (due to clogging) and models batch-fed, daily at 0.7 l/d, 75% PCS
77	Feed strength increased to 100% PCS, with 3 g/l $NaHCO^3$ (N & P still added to give COD:N:P = 500:5:1) and 0.5 g/l yeast extract
92	Increased $NaHCO_3$ to 4 g/l
99	Daily therapeutic doses of 15 ml/d of sludge started
115	Last day of therapeutic sludge addition
127	External mixing device installed to help improve gas collection. Mixer squeezed riser tube to disturb model contents to help relieve entrapment of gas bubbles due to fibrous nature of sludge
140	Yeast extract deleted from feed (out of stock)
147	$NaHCO_3$ deleted and N added as NH_4HCO_3 (4.5 g/l)
150	Wood media removed and modified to increase pore space to reduce clogging

TABLE 7.4 PCS DFF NO. 1 & NO. 2 OPERATION (Continued)

Day	Event/Comments
154	Yeast extract reintroduced to feed (0.5 g/l)
155	Plexiglas covers removed and replaced with plastic funnels to reduce headspace
182*	Feed rate changed from 0.7 l/d to 0.3 l/d to increase HRT and improve digestion. Therapeutic sludge addition of 15 ml/d for next 40 days. $NaHCO_3$ reintroduced (4 g/l)
189	$NaHCO_3$ decreased to 3 g/l
203	Model mixer removed - not working well. Contents of reactor mixed daily by raising and lowering media in model - continued to end of study.
210	$NaHCO_3$ decreased to 2 g/l
211	Plexiglas cover replaced on model since funnel covers made no improvement in gas collection
241	Gas collection by displacement of model contents begun (Figure 6.5)
252	Rumen fluid added to PCS DFF No. 2 only (35°C model) 100 ml/d for next 6 days to improve digestion
276	Model shut down

* day 182 to end called "steady-state" period in report.

Throughout the study chemicals were added to both the HYSSL and PCS to provide nutrients, alkalinity and pH control. N and P were added as NH_4Cl and KH_2PO_4, respectively, in quantities sufficient to ensure at least a COD:N:P ratio of approximately 500:5:1. For a brief period N was added as NH_4HCO_3. In addition, yeast extract (0.5 g/l) was added to both wastes to supply adequate micronutrients. To supplement alkalinity and provide pH control, various concentrations (from 1 to 4 g/l) of $NaHCO_3$ (NH_4HCO_3 was used briefly) were added to the feeds. For short periods KOH (or $Ca(OH)_2$) was added to HYSSL feeds to provide pH control. More detailed descriptions of the steady-state operations are presented in Tables 7.1 through 7.4.

7.3 Analyses

In order to monitor model performance, daily grab samples of feed and effluent were collected from each reactor and combined to form weekly composites. The weekly composites from the HYSSL models were analyzed for COD, BOD, SS and VSS. As well, twice weekly, grab samples of HYSSL model feed and effluent were analyzed for pH, VA and alkalinity. Once per week grab samples of feed and effluent were measured for total sulphide.

The weekly composites of PCS model feed and effluent were tested for COD, TS and TVS; twice weekly grab samples were tested for pH, VA and alkalinity. The COD, TS and TVS results are not reported here due to contradictory performance (i.e., COD, TS and TVS removals) results, arising from sampling difficulties posed by the fibrous and clumping nature of the PCS (details Section 9.1).

Biogas production was measured from each model (as described in Section 6.4) and weekly samples were analyzed for $CO_2(+H_2S)$ and O_2, using an ORSAT apparatus, and H_2S separately. Methane component was determined by difference. The quantities of gases other than CO_2 and H_2S, i.e., H_2, N_2 and other trace gases were assumed to be negligible.

Periodically, samples of PCS and HYSSL raw wastes were subjected to more complete analysis. The results of these analyses are summarized in Tables 4.1 and 4.2.

A complete list of analytical tests and procedures used in this study are listed in Table 7.5.

TABLE 7.5 ANALYTICAL PROCEDURES

Parameter	Procedure
BOD	Standard methods (20)
COD	Dichromate digestion, Hach
SS	Standard Methods
VSS	" "
TS	" "
TVS	" "
TKN	" "
NH_3-N	" "
Total P	" "
Total S	TAPPI Standard Method (21)
SO_3	Standard Methods
SO_4	" "
Sulphide	" "
Alkalinity	" "
VA	See Reference 22
Ca	Standard Methods
Cd	Atomic Absorption (AA)
Cr	AA
Cu	AA
Fe	AA
K	AA
Mg	AA
Na	AA
Ne	AA
Pb	AA
Zn	AA
Hg	Standard Methods
Gas - $CO_2(+H_2S)+ O_2$	ORSAT
Gas - H_2S	Tutweiler method (23), MSA apparatus (Samplair Pump and H_2S Tubes)

8.0 HYSSL RESULTS

8.1 ADI-BVF

The temporal variations of BOD (and total sulphide), COD, SS, VA (and alkalinity) and pH are illustrated in Figures 8.1, 8.2, 8.3, 8.4 and 8.5, respectively. The BOD, COD and SS values shown in these figures represent the results of tests on weekly composites while total sulphide, VA, alkalinity and pH are representative of grab samples taken twice weekly.

Average values for BOD, COD, SS, total sulphide, VA, alkalinity and pH over the period of "steady-state" operation (from day 182 onwards) are presented in Table 8.1, along with average percent removals of BOD, COD and SS. Results of total sulphur tests are given in Section 8.5.

TABLE 8.1 HYSSL BVF DATA SUMMARY

Parameter	Influent	Effluent	Removal (%)
BOD	4 420	1 180	73.3
COD	26 000	17 100	34.2
SS	1 050	190	81.9
Total Sulphide	330	280	-
VA	1 640	1 360	-
Alkalinity	900	3 100	-
pH	5.9	8.0	-

During the "steady-state" period the average BOD and COD loads were 0.42 and 2.48 $kg/m^3.d$, respectively. Average HRT was approximately 10.3 d. The specific CH_4 production rate was 0.52 m^3 CH_4/kg BOD removed (0.25 m^3 CH_4/kg COD removed). Biogas composition averaged approximately 51% CH_4, 40.5% CO_2 and 8.5% H_2S. Variations of BOD load and HRT are illustrated in Figure 8.6.

FIGURE 8.1: HYSSL BVF - TOTAL SULPHIDE AND BOD

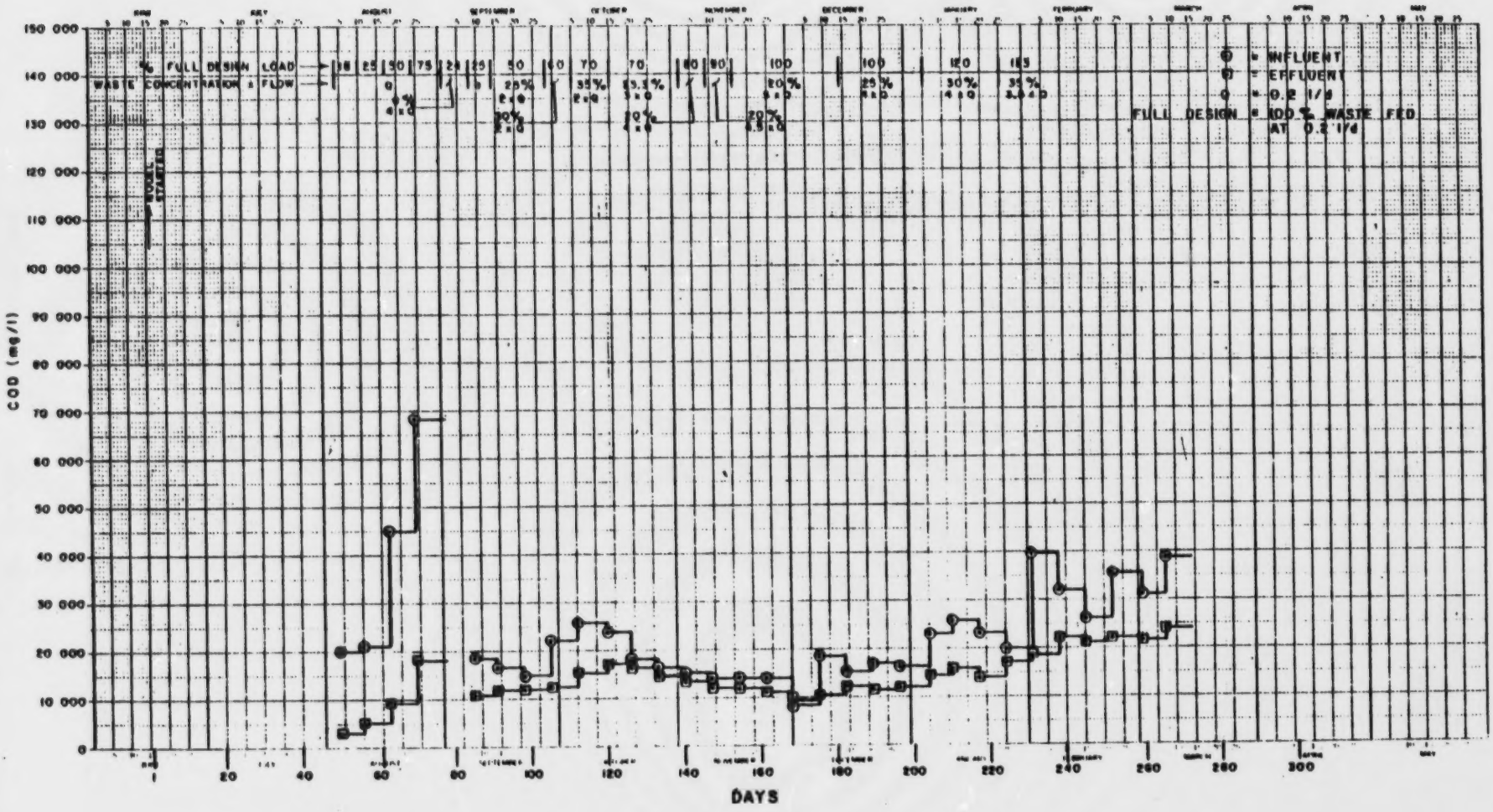


FIGURE 8.2: HYSSL BVF – COD

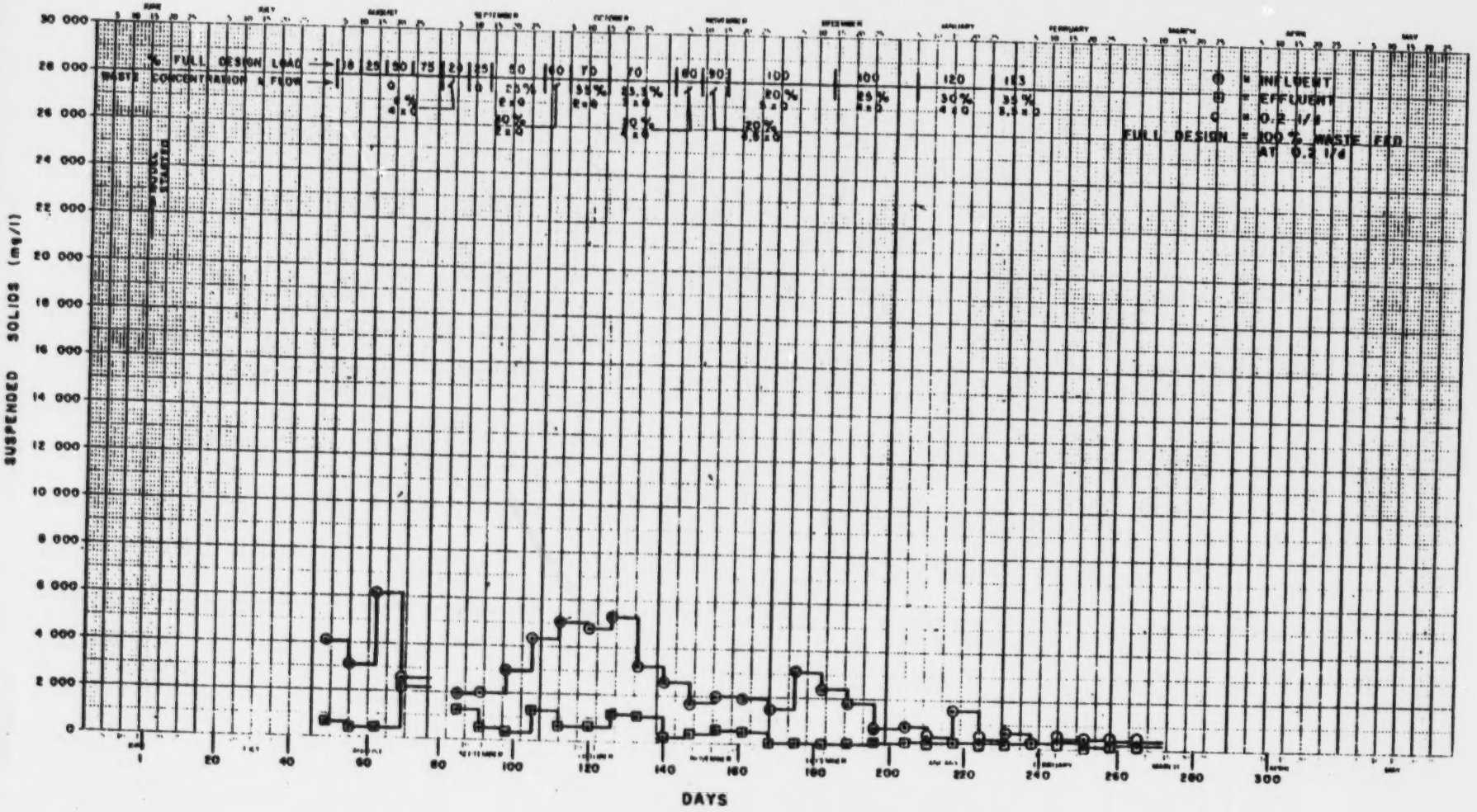


FIGURE 8.3: HYSSL BVF – SS

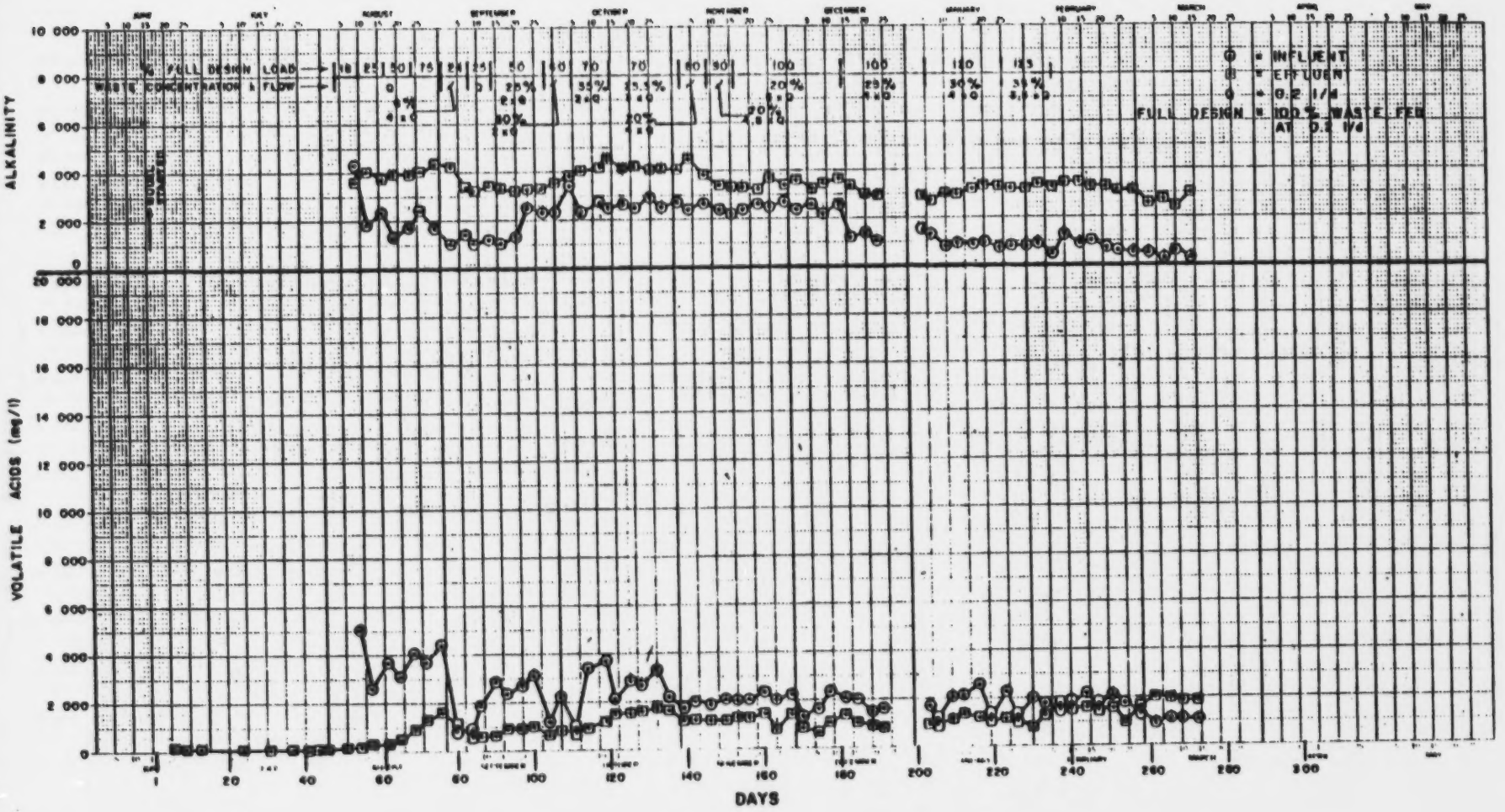


FIGURE 8.4: HYSSL BVF — VA AND ALKALINITY

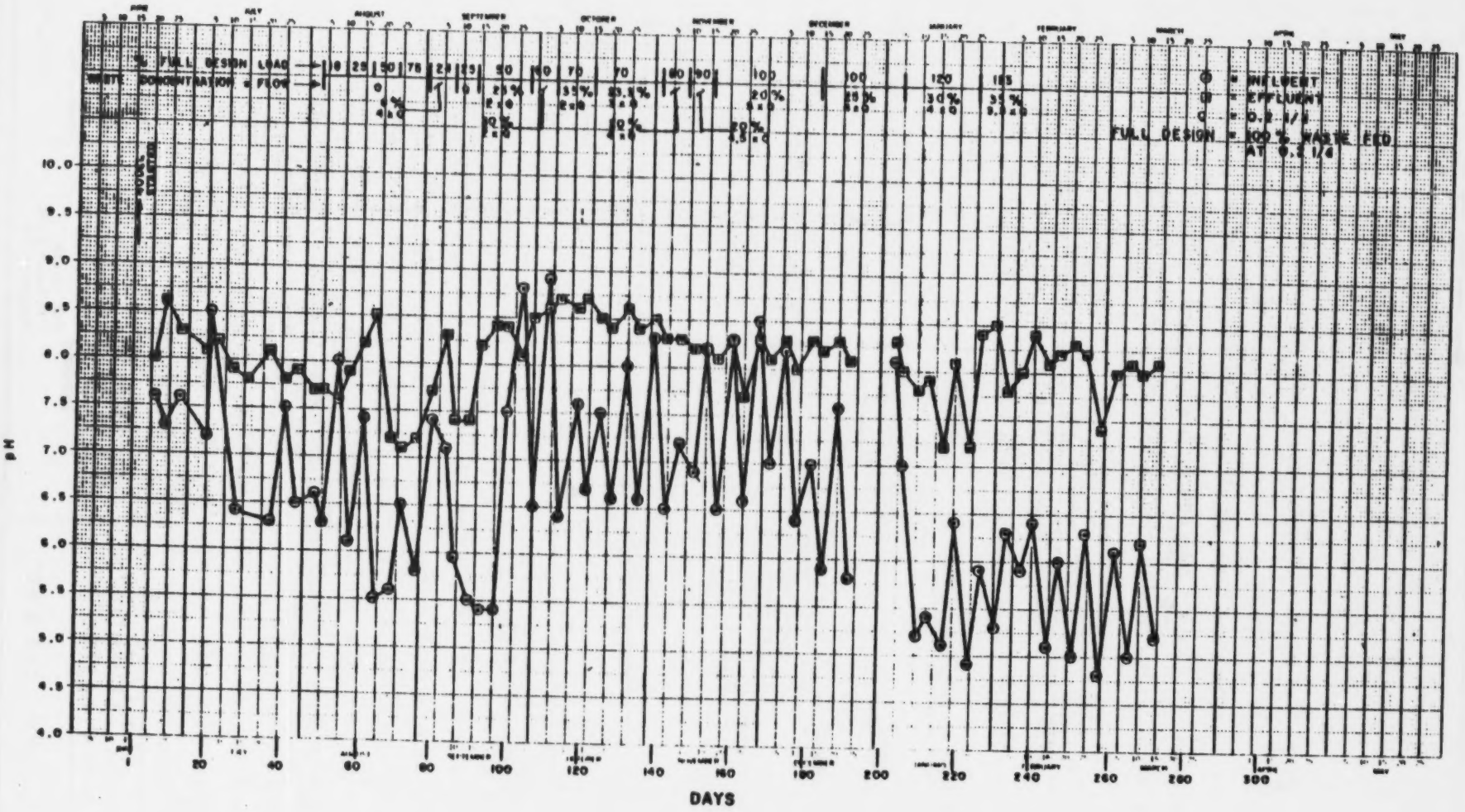
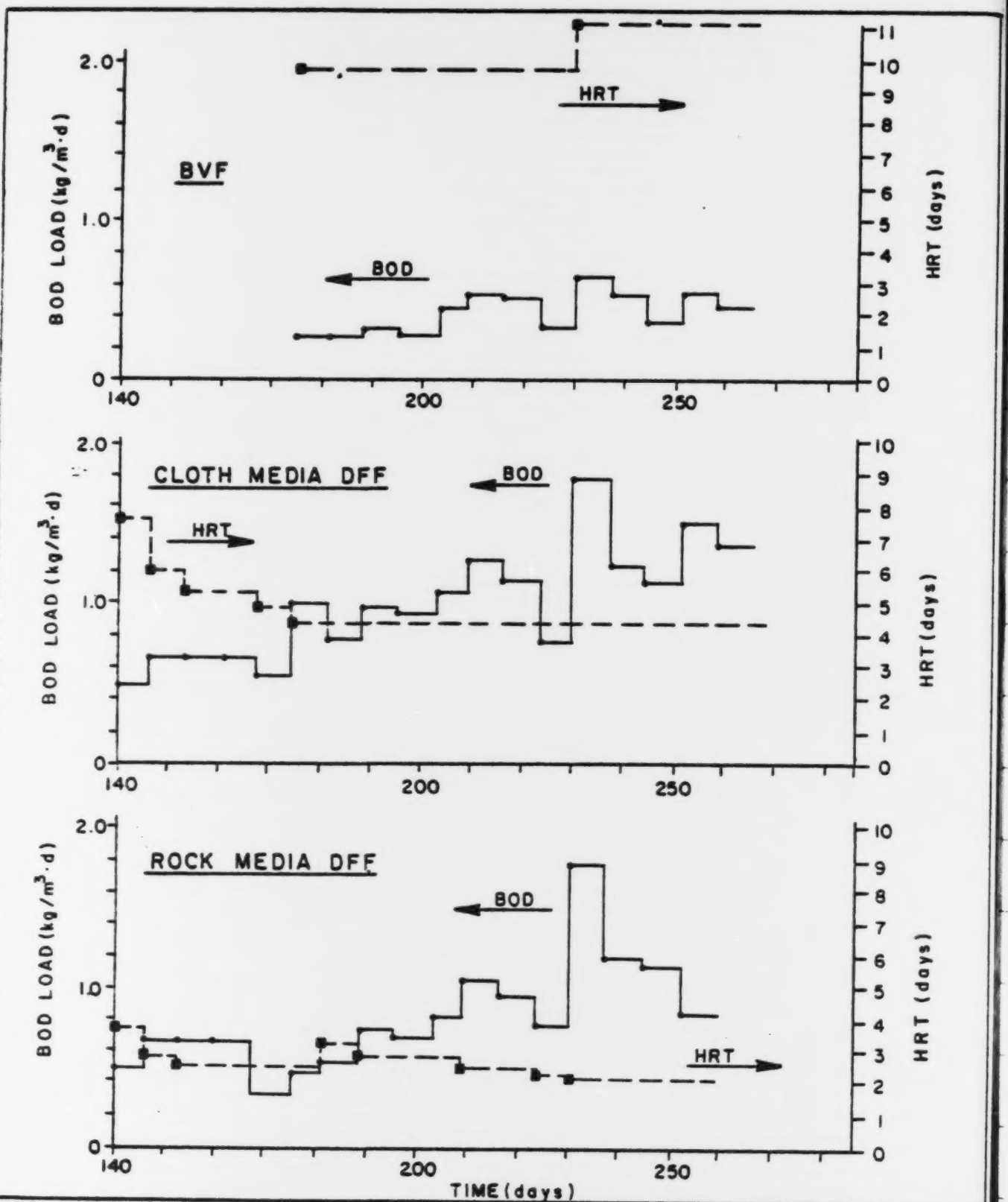


FIGURE 8.5: HYSSL BVF – pH



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Drawing: HYSSL MODELS-BOD LOAD & HRT		Drawing No.: FIGURE 8.6		

8.2 Cloth Media DFF

The temporal variations of BOD (and total sulphide), COD, SS, VA (and alkalinity) and pH are presented in Figures 8.7, 8.8, 8.9, 8.10 and 8.11, respectively. BOD, COD and SS values are representative of weekly composites and the total sulphide, VA, alkalinity and pH are from grab samples taken twice weekly.

Average values for BOD, COD, SS, total sulphide, VA, alkalinity and pH over the period of "steady-state" operation (from day 147 onwards) are presented in Table 8.2, along with average BOD, COD and SS removals.

TABLE 8.2 HYSSL CLOTH DFF DATA SUMMARY

Parameter	Influent	Effluent	Removal (%)
BOD	5 180	1 500	71.0
COD	30 100	21 100	29.9
SS	1 710	2 990	(increase)
Total Sulphide	390	350	-
VA	2 230	1 600	-
Alkalinity	1 390	3 350	-
pH	6.1	7.9	-

During the steady-state period average BOD and COD load and HRT were 1.00 $kg/m^3.d$, 5.73 $kg/m^3.d$ and 4.9 d, respectively. The specific CH_4 production was 0.56 m^3 CH_4/kg BOD removed (0.26 m^3 CH_4/kg COD removed). Biogas composition averaged approximately 69% CH_4, 27.1% CO_2 and 3.9% H_2S. The biogas composition for this model is markedly different from the others as it was scrubbed with caustic solution to remove CO_2 and H_2S from day 219 on. This gas was then recycled through the reactor to strip H_2S. Figure 8.6 presents the BOD load and HRT variation. Results of dynamic testing (feed cessation and shock load) are presented in Section 8.4 and Section 8.5 gives results of total sulphur tests.

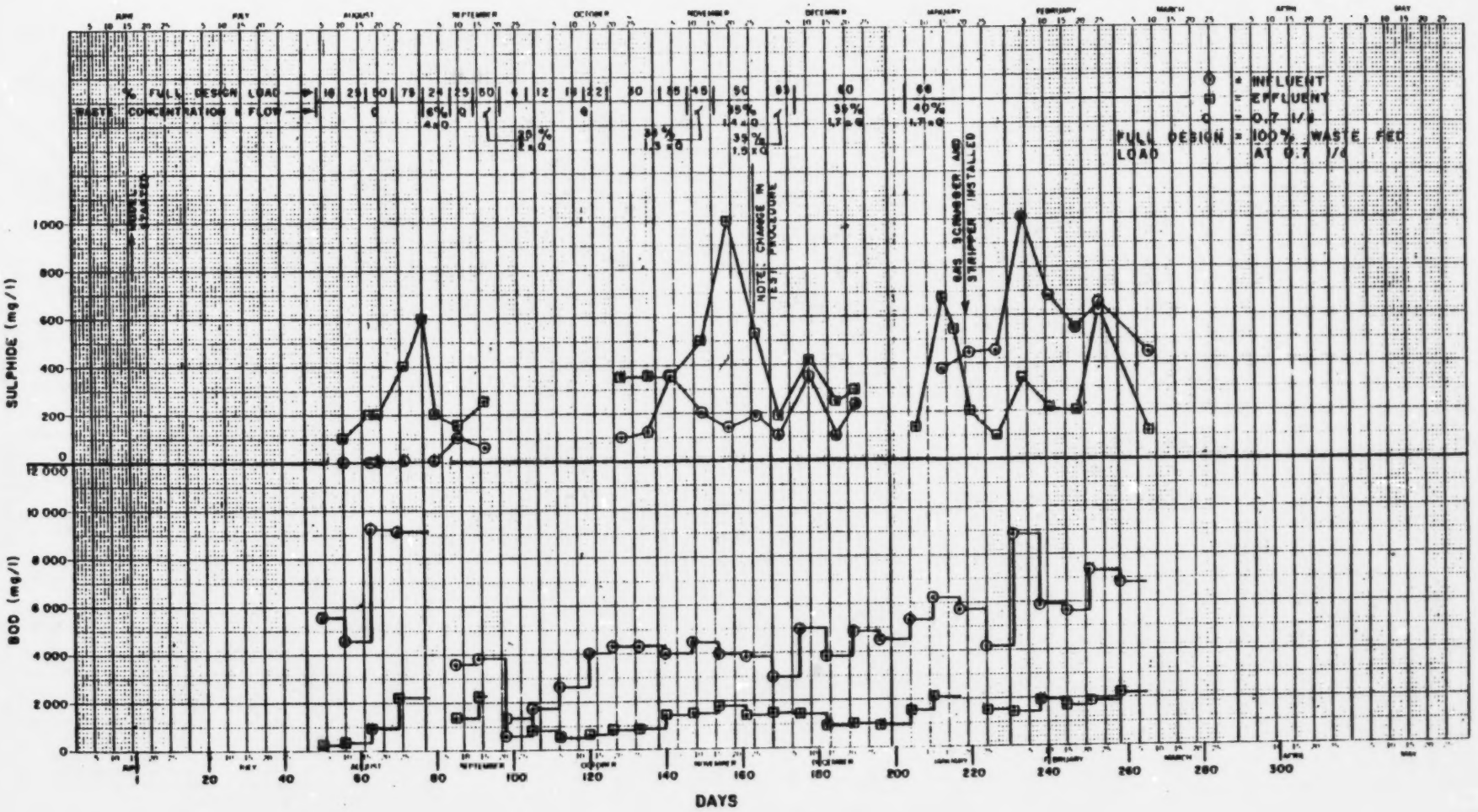


FIGURE 8.7: HYSSL CLOTH MEDIA DFF — TOTAL SULPHIDE AND BOD

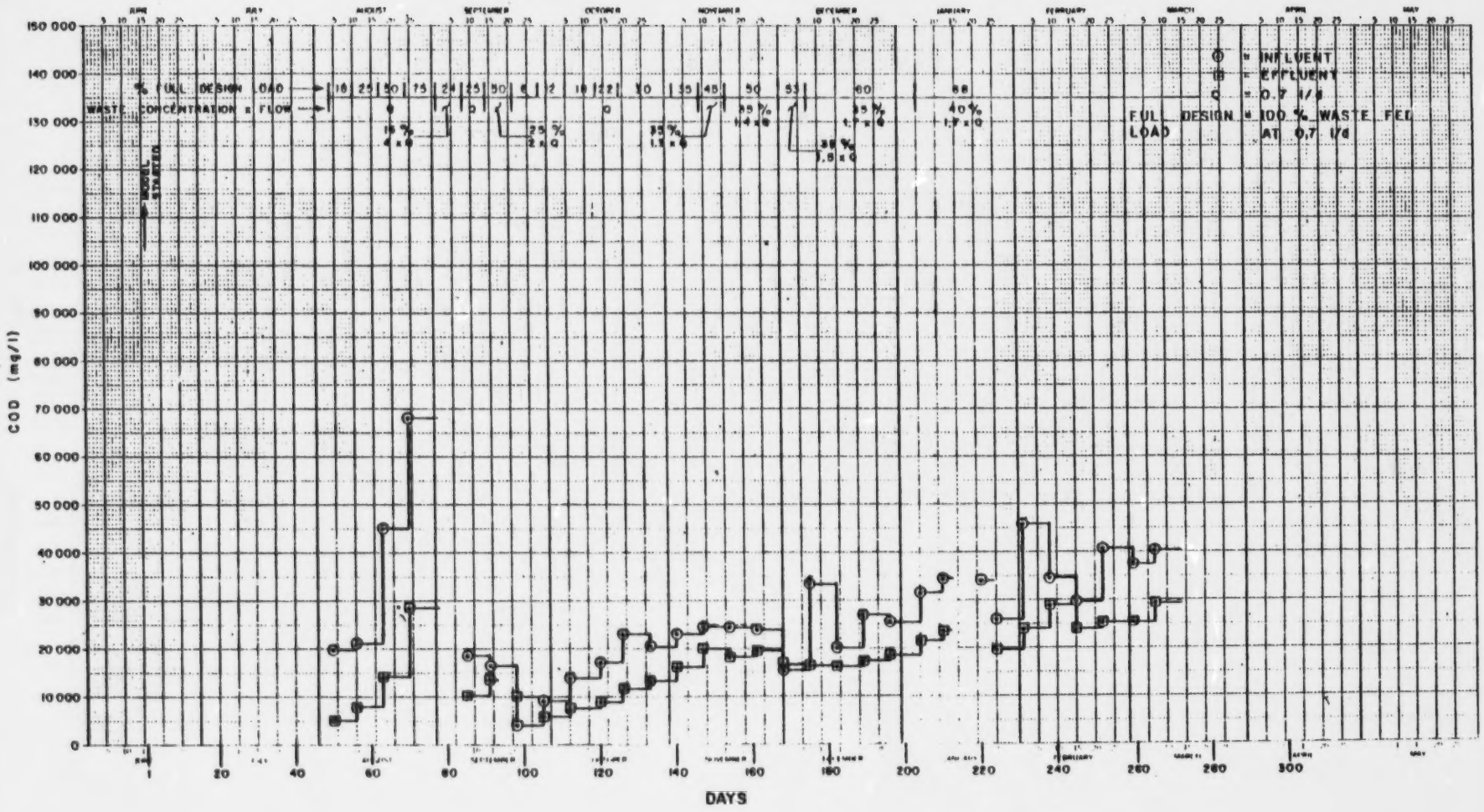


FIGURE 8.8: HYSSL CLOTH MEDIA DFF — COD

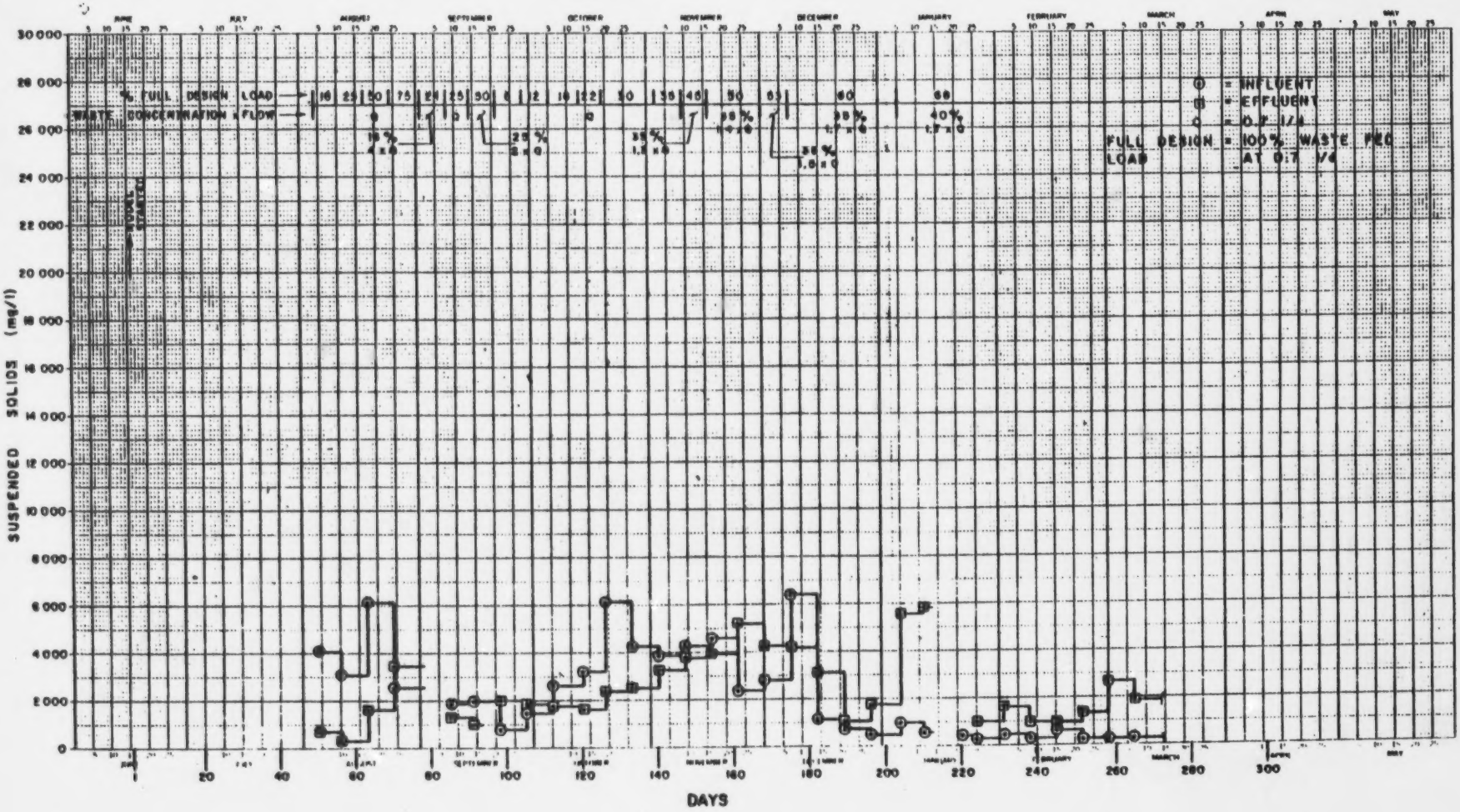


FIGURE 8.9: CLOTH MEDIA DFF – SS

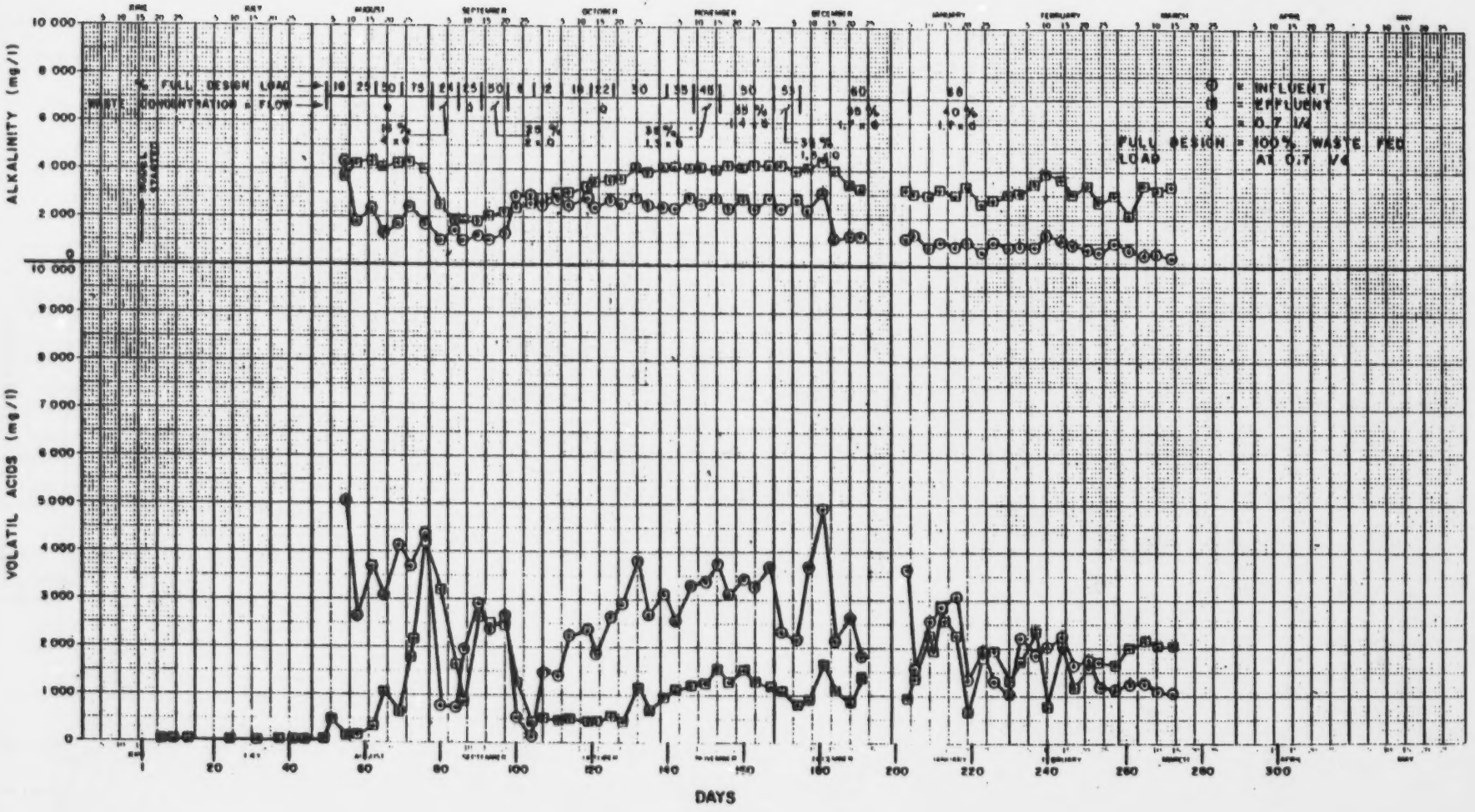


FIGURE 8.10: HYSSL CLOTH MEDIA DFF – VA AND ALKALINITY

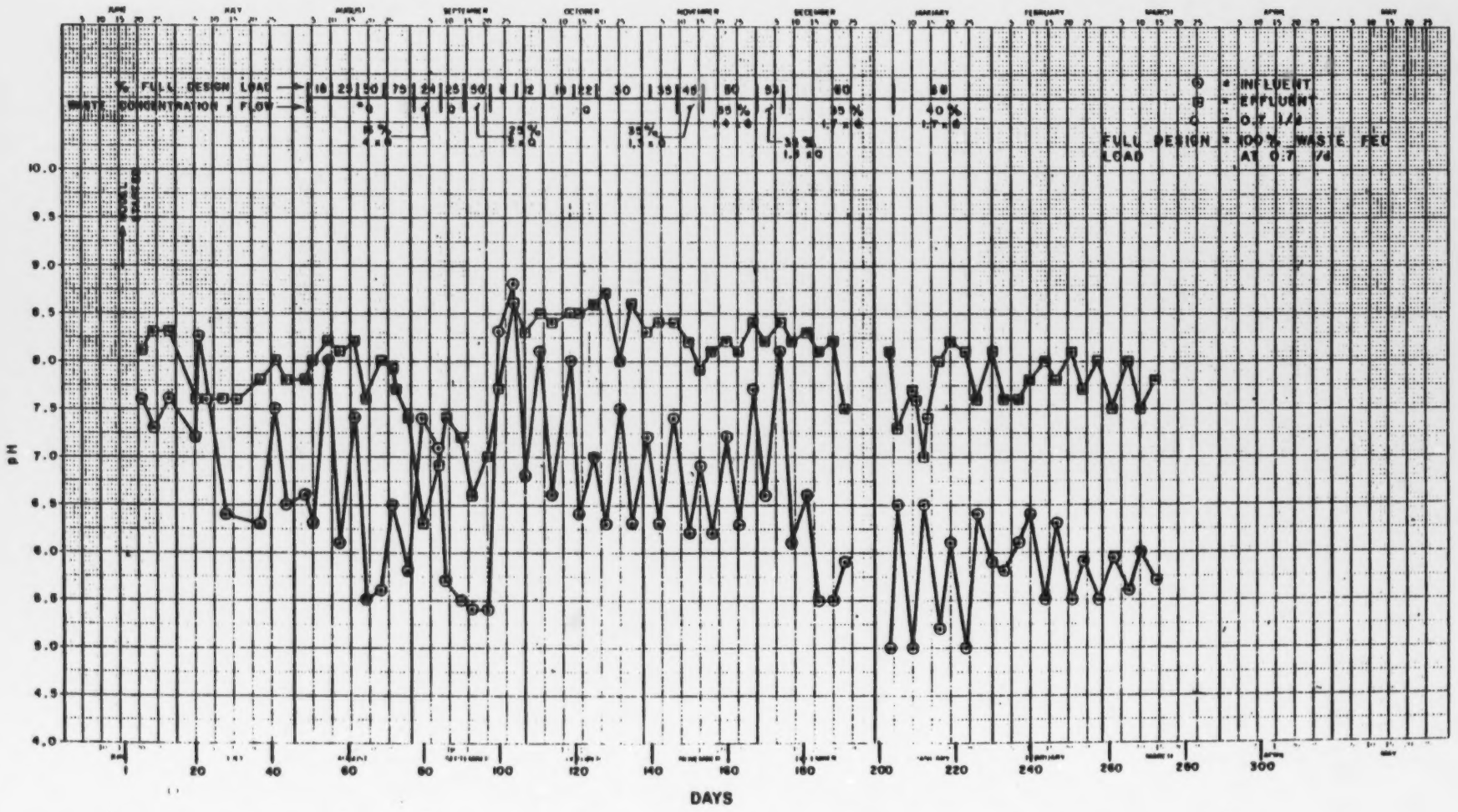


FIGURE 8.11: HYSSL CLOTH MEDIA DFF — pH

8.3 Rock Media DFF

Figures 8.12, 8.13, 8.14, 8.15 and 8.16 represent the variation of BOD (and total sulphide), COD, SS, VA (and alkalinity) and pH, respectively, with respect to time. Total sulphide, VA, alkalinity and pH results represent analyses of two grab samples per week and BOD, COD and SS results represent weekly composites.

Average values for BOD, COD, SS, total sulphide, VA, alkalinity and pH from day 147 onwards (steady-state operation) are shown in Table 8.3, along with average BOD, COD and SS removals.

TABLE 8.3 HYSSL ROCK DFF DATA SUMMARY

Parameter	Influent	Effluent	Removal (%)
BOD	4 670	1 530	67.2
COD	26 700	17 900	33.0
SS	1 600	1 480	7.5
Total Sulphide	360	320	-
VA	2 160	1 290	-
Alkalinity	1 570	3 320	-
pH	6.2	7.9	-

During steady-state operation BOD load, COD load and HRT averaged 0.81 kg/m^3.d, 4.61 kg/m^3.d and 2.7 d, respectively. The specific CH_4 production averaged 0.47 m^3 CH_4/kg BOD removed (0.21 m^3 CH_4/kg COD removed). Biogas composition averaged approximately 49% CH_4, 42.2% CO_2 and 8.8% H_2S. Figure 8.6 presents the temporal variation of BOD load and HRT.

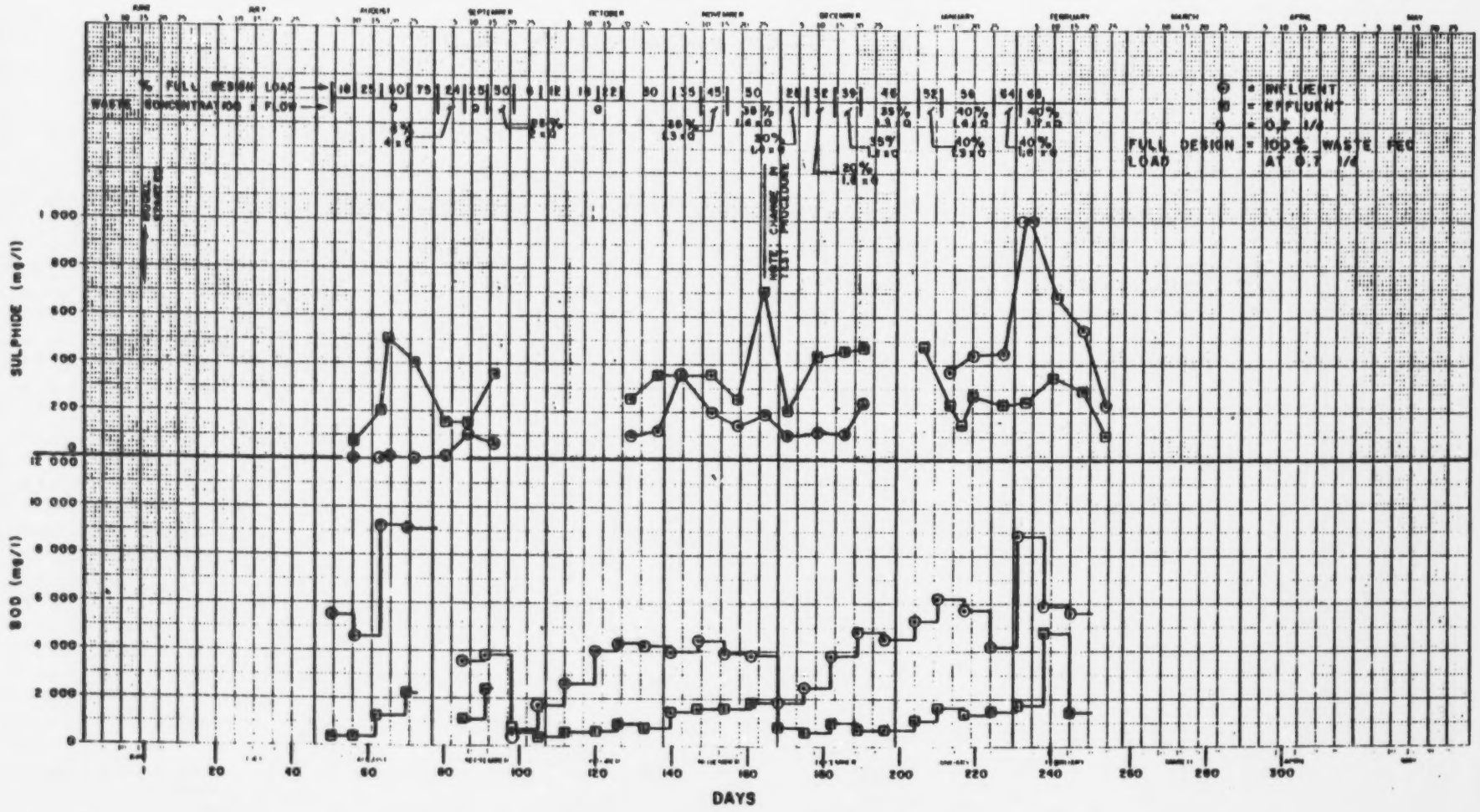

% FULL DESIGN LOAD →
WASTE CONCENTRATION x FLOW →
SULPHIDE (mg/l)
BOD (mg/l)
DAYS
WATER CHARGE IN TEST PROCEDURE
FULL DESIGN LOAD = 100% WASTE REC AT 0.7 l/d

FIGURE 8.12: HYSSL ROCK MEDIA DFF — TOTAL SULPHIDE AND BOD

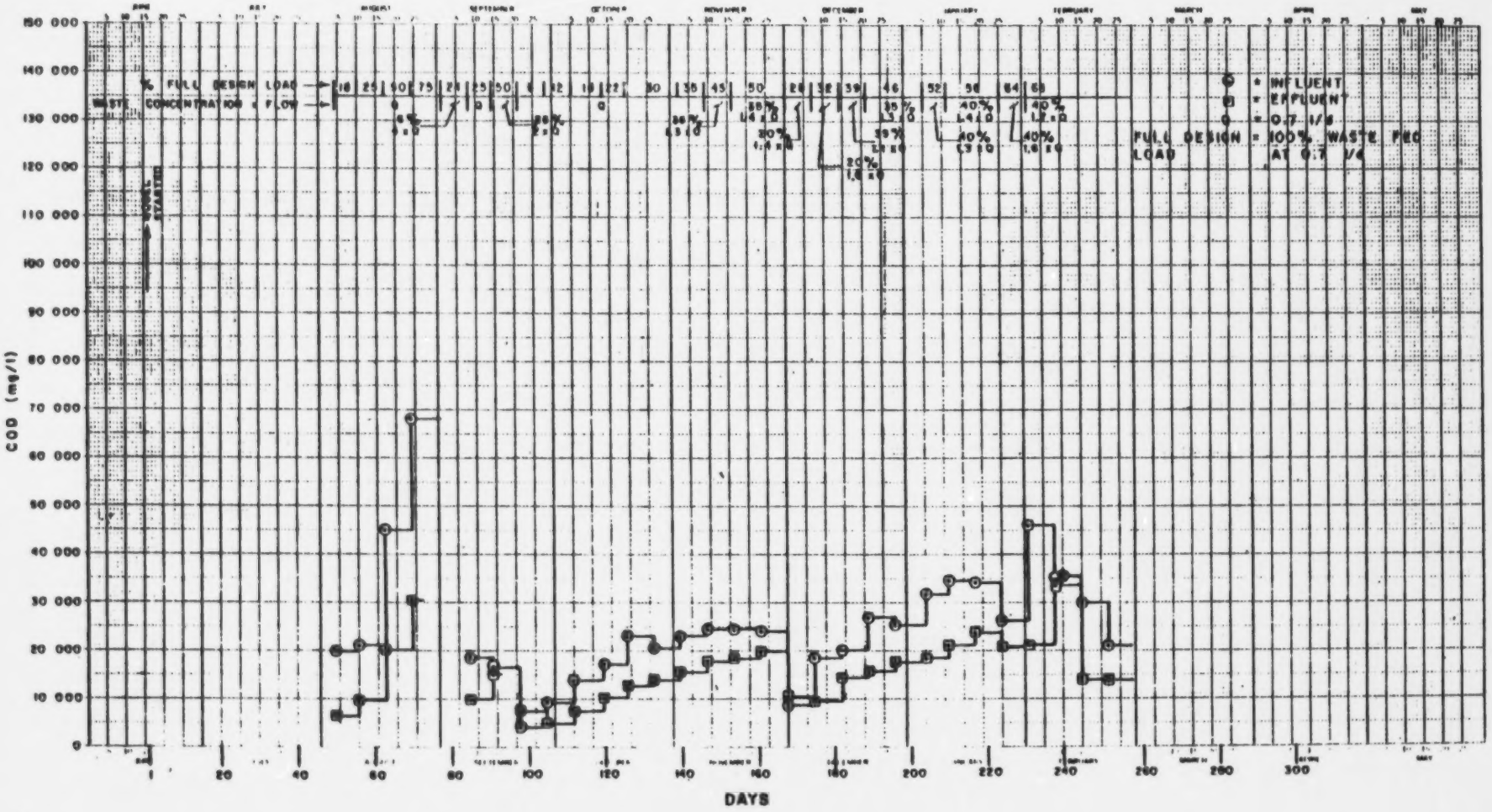


FIGURE 8.13: HYSSL ROCK MEDIA DFF – COD

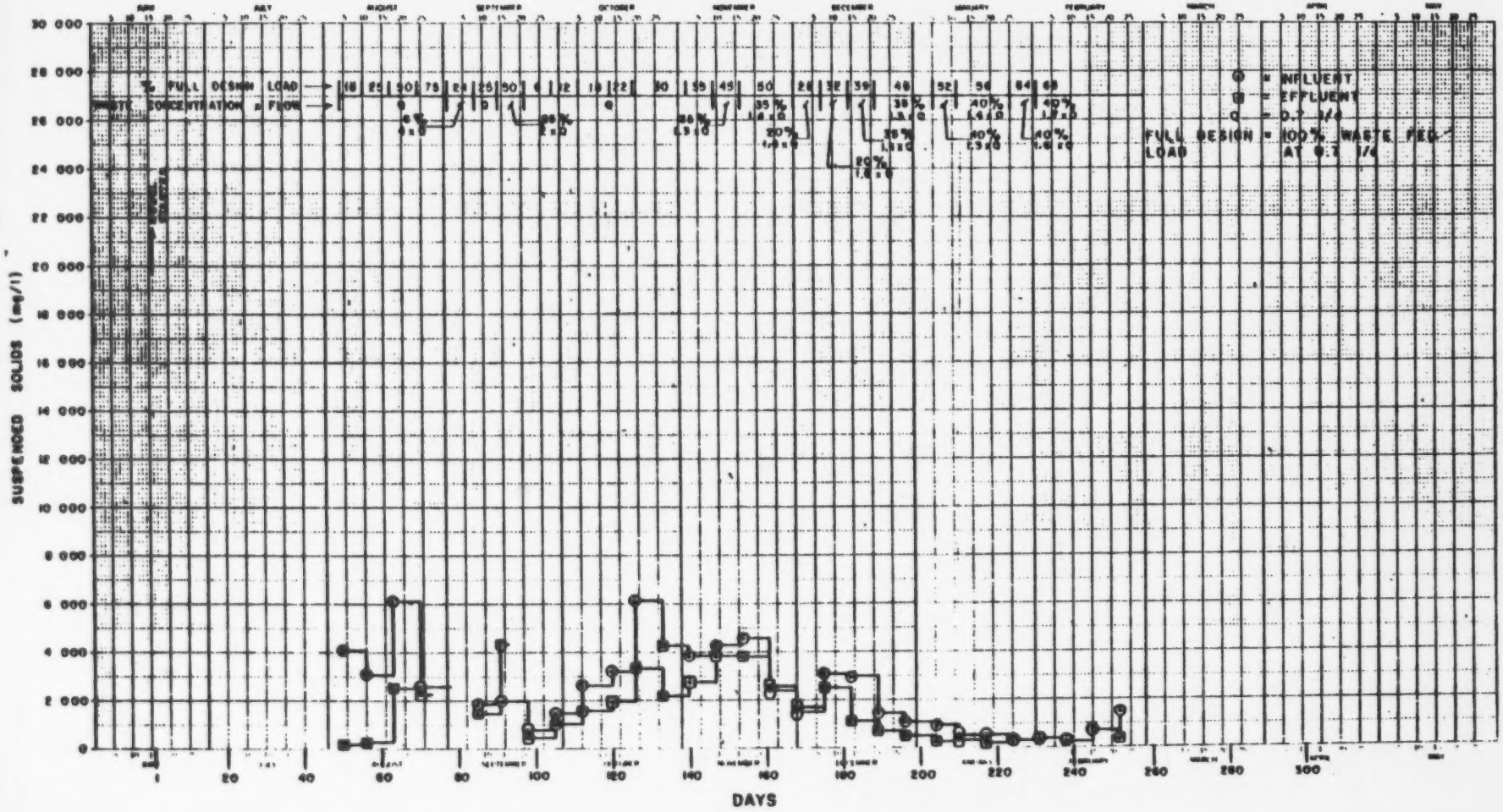


FIGURE 8.14: HYSSL ROCK MEDIA DFF - SS

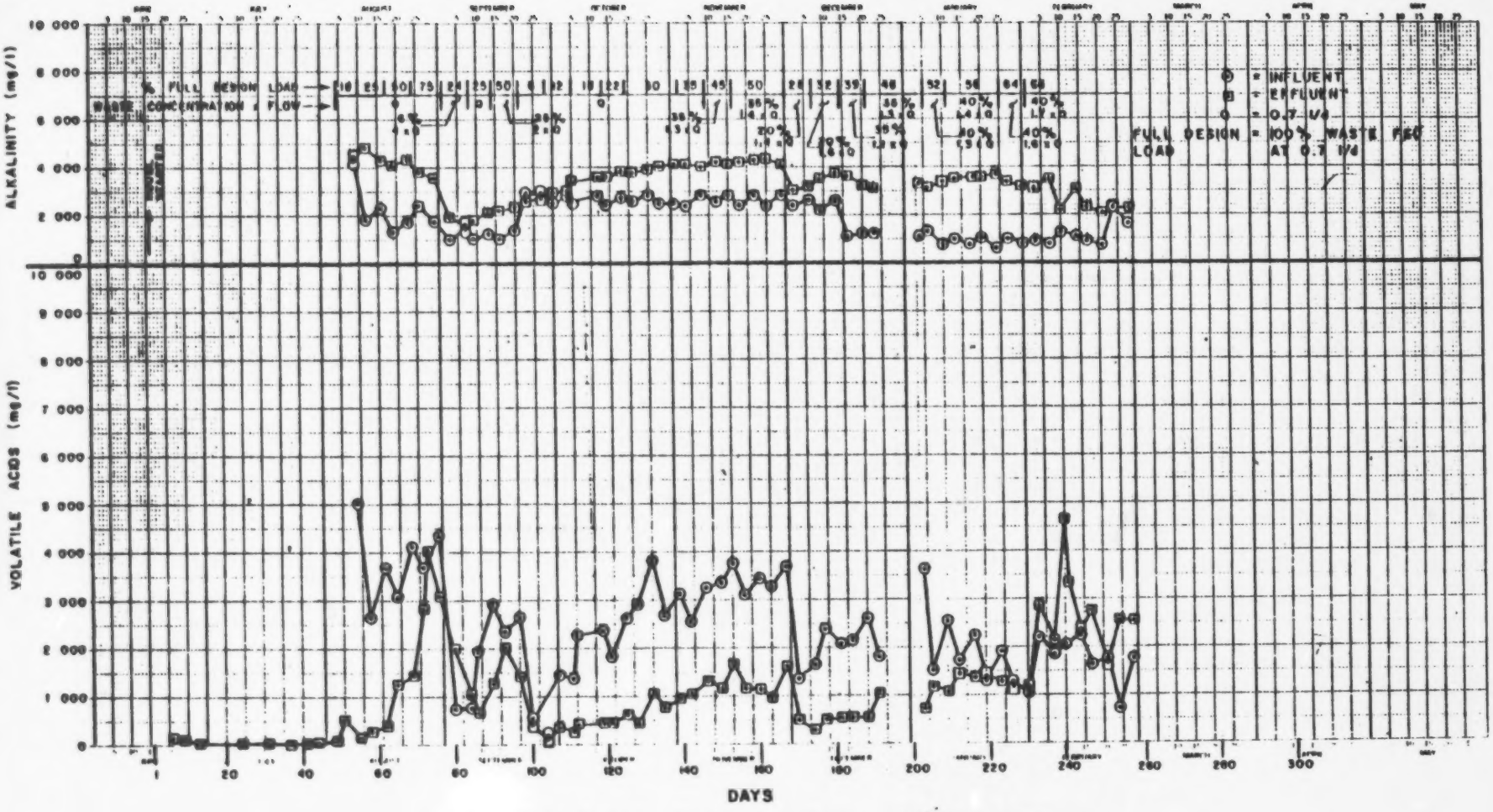


FIGURE 8.15: HYSSL ROCK MEDIA DFF — VA AND ALKALINITY

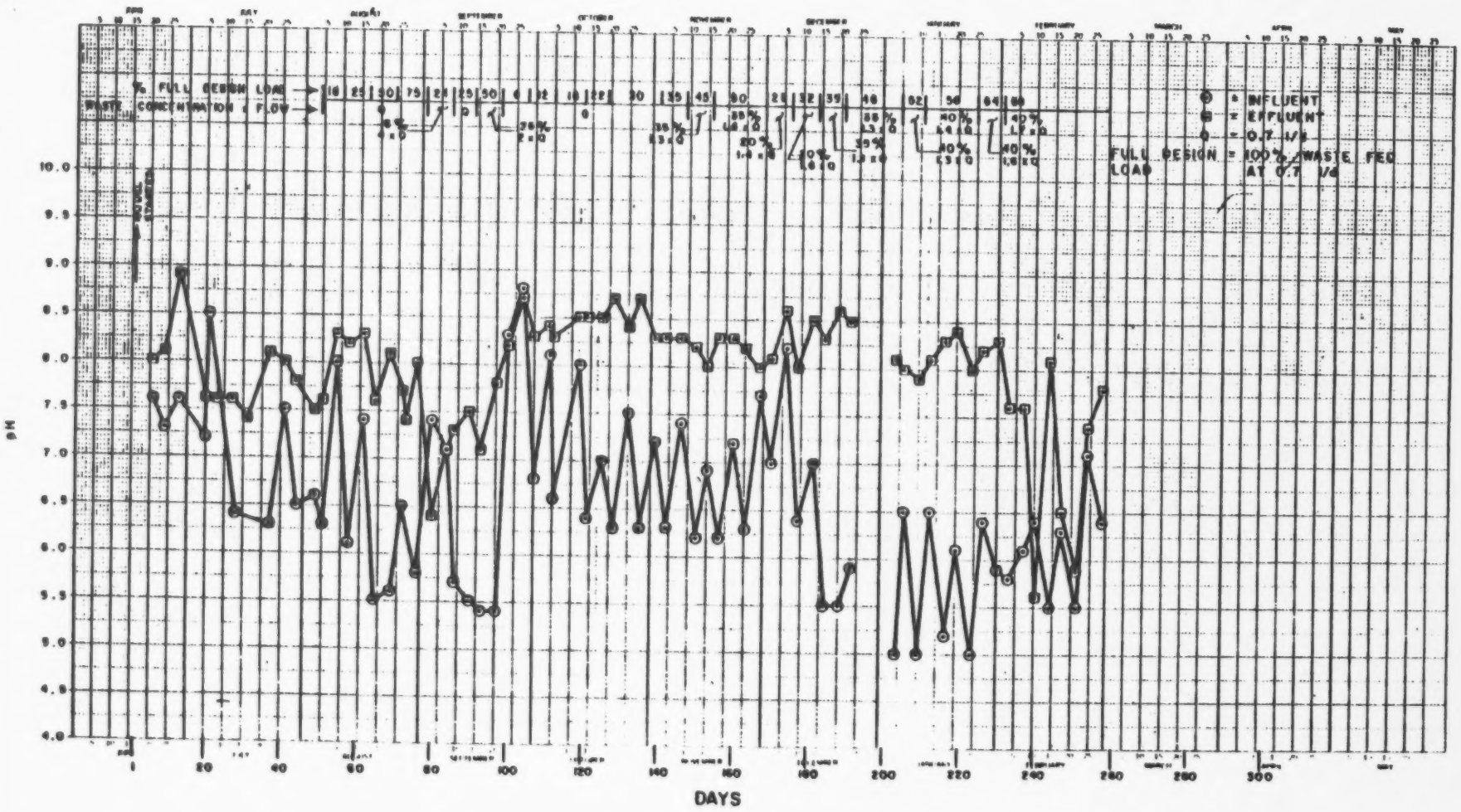


FIGURE 8.16: HYSSL ROCK MEDIA DFF – pH

8.4 Dynamic Testing

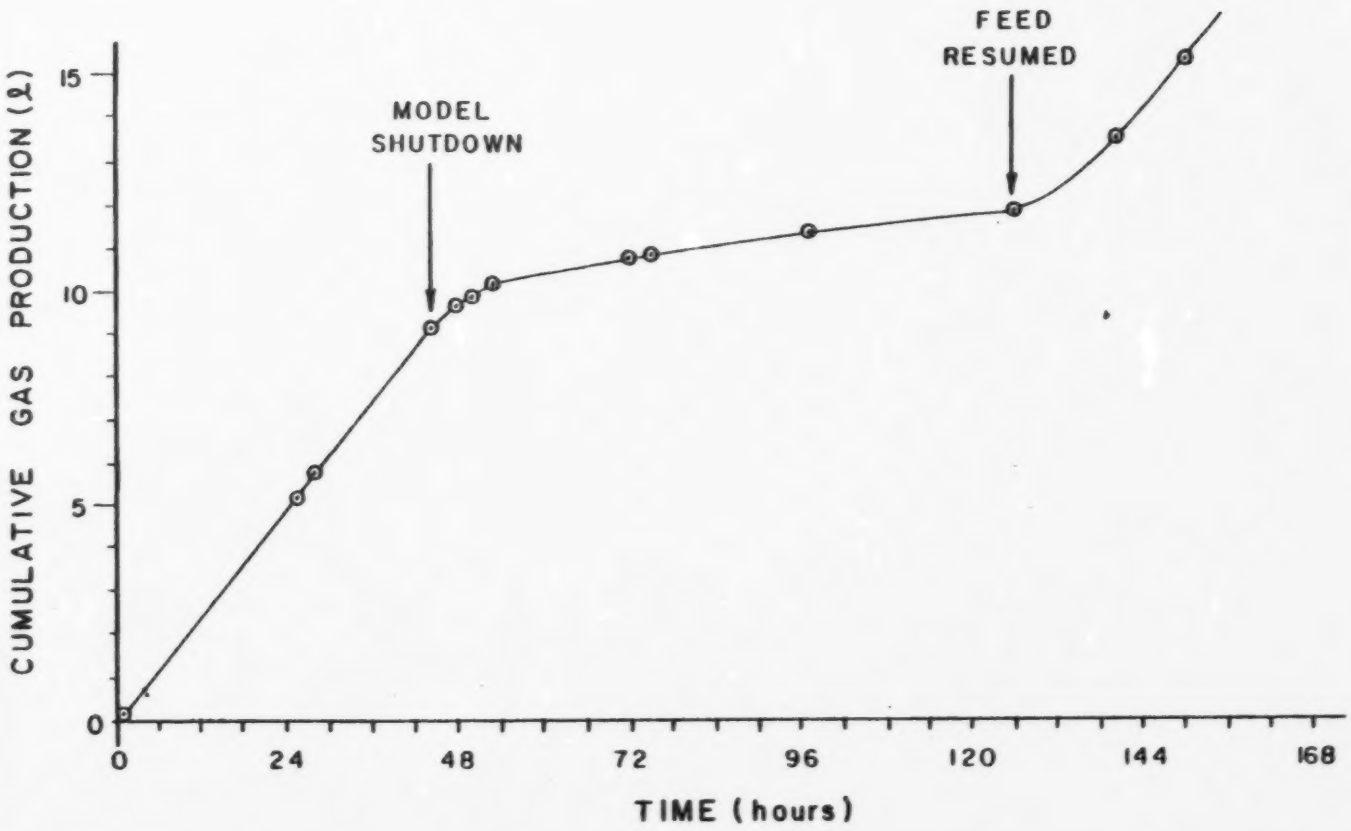
In order to simulate what could occur in practice, two series of dynamic tests were carried out on the HYSSL cloth media DFF. The first tests monitored the response (gas production) of the model to a cessation (of approximately 3.5 days duration) and subsequent resumption of feed. The second measured CH_4 production and VA levels resulting from a 24 hour shock load of three times the normal flow. The cloth media DFF was selected for dynamic testing because it was stable at the time of testing while the rock media DFF was under stress from what was believed to be sulphide inhibition/toxicity.

Figure 8.17 illustrates the biogas production response to shutdown and resumption of feed. At the time of this test the model HRT and BOD loads were 4.5 days and 1.43 $kg/m^3.d$, respectively.

Figures 8.18 and 8.19 show the change in CH_4 production and VA levels as a result of a 24 hour shock load (flow rate tripled). The increased flow rate decreased the reactor HRT to 1.5 days and increased the BOD load to 4.30 $kg/m^3.d$.

8.5 Sulphur Balance

In order to determine the sulphur mass balance through the anaerobic system, the total sulphur (TAPPI standard method) concentration of BVF and cloth media DFF model influent and effluent was measured. Total sulphur analyses were carried out on weekly composites for three weeks on the BVF (day 252 to 272) and two weeks (day 259 to 272) on the cloth media DFF. To complete the sulphur picture the sulphide content of the biogas streams was also recorded. In the case of the DFF only a small



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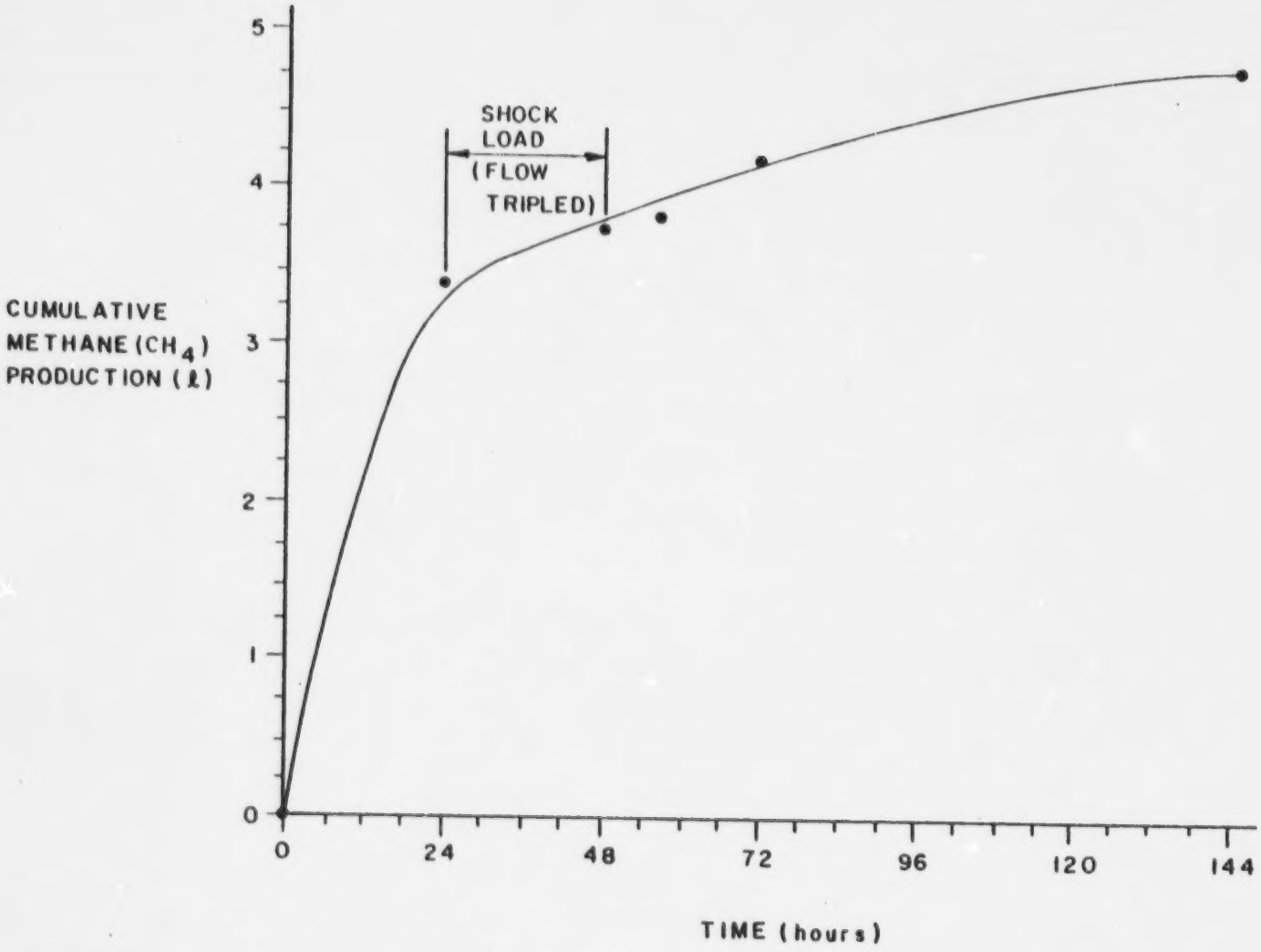
Drawing: BIOGAS PRODUCTION RESPONSE TO HYSSL FEED CESSATION		Drawing No.: FIGURE 8.17		

ADI Z-107



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Drawing: CLOTH MEDIA DFF METHANE PRODUCTION RESPONSE TO SHOCK LOAD		Drawing No.: FIGURE 8.18		

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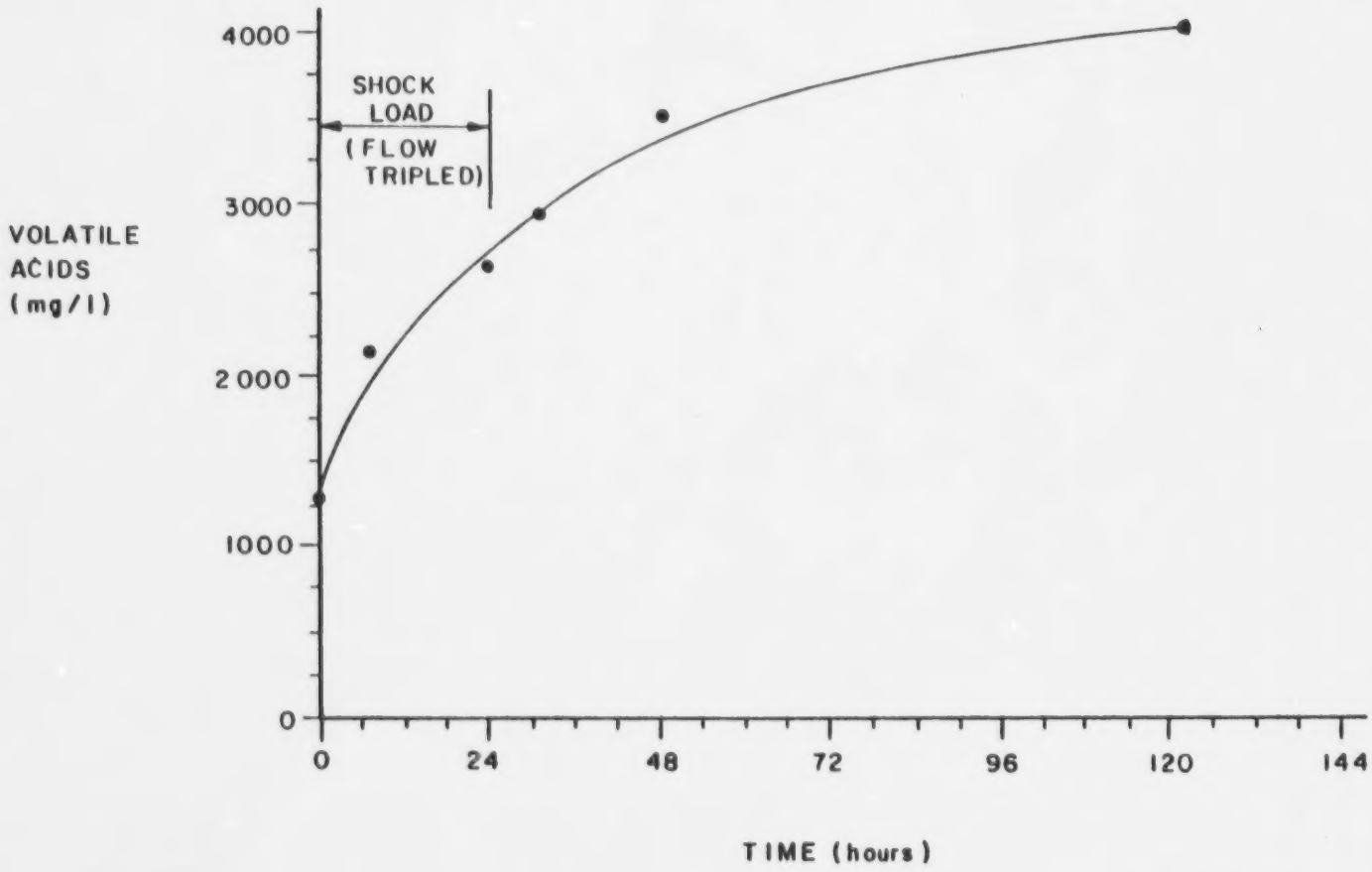

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Drawing: CLOTH MEDIA DFF VA RESPONSE TO SHOCK LOAD		Drawing No. FIGURE 8.19		

ADI 2-157

portion of the sulphur in the biogas was actually determined; the sulphur removed from the biogas by the scrubber-stripper was not measured. Tables 8.4 and 8.5 present the total sulphur, biogas sulphide, liquid flow and biogas production data collected for the BVF and cloth media DFF, respectively.

TABLE 8.4 BVF SULPHUR DATA

	Total Sulphur (mg/l)		Feed Rate	Biogas H_2S	Biogas Production
Week	Influent	Effluent	l/d	(Vol %)	(l)
1	2 820	1 890	0.7	9.3	28.6
2	2 600	960	0.7	10.2	32.9
3	1 750	1 160	0.7	7.5	34.4

TABLE 8.5 CLOTH MEDIA DFF SULPHUR DATA

	Total Sulphur (mg/l)		Feed Rate	Biogas H_2S	Biogas Production
Week	Influent	Effluent	l/d	(Vol %)	(l)
1	2 120	540	1.2	1.5	29.7
2*	2 420	1 150	1.2	0.3	13.5

*Cloth model fed for only 3.5 days during week 2 due to feed cessation phase of dynamic testing in progress.

Using the flow and concentration data, presented in Tables 8.4 and 8.5, the total sulphur entering (influent) and leaving (effluent + biogas) each model was calculated. Over a three week period 35.1 g of sulphur was fed into the BVF, 19.7 g left in the effluent and 12.3 g left in the biogas. Therefore, over the test period, of the 35.1 g of sulphur fed to the model, 32.0 g of sulphur could be accounted for in the biogas and effluent.

Over a two week period 28.0 g of sulphur was fed to the cloth media DFF, 9.4 g left in the effluent and it is reasoned that most of the remainder (i.e. 18.6 g) left in the biogas. Of this amount, the scrubber-stripper apparently removed all but 0.7 g which was measured in the biogas collected. These results suggest the scrubber-stripper (as crude as it was), was fairly effective in removing sulfur from the reactor contents and the biogas as well. It is also interesting to note that the sulphide stripping occurred at relatively high pH. If a proper stripper had been employed and a lower model pH maintained the sulphur removals would have been higher. This concept also opens up the possibility of sulphur recovery.

The sulphur balance results for each model are discussed in greater detail in Section 10.3

9.0 PCS RESULTS

9.1 Operational Problems

There were numerous operational problems with both PCS DFF models, mostly related to the fibrous nature of the sludge (and its large solids particles). During startup dilute PCS was pumped to the models with peristaltic pumps, but as waste concentration increased this proved difficult due to frequent clogging. Therefore, from day 76 onwards the models were batchfed, by hand, once daily. Some of the packing was also removed from the original wood media to increase pore size and reduce clogging.

After startup, gas collection also proved difficult and only a very small percentage of the biogas produced was collected. Most of the gas was trapped by sludge particles and clumps (many of which also clogged the media).

A mechanical mixing system which exerted a squeezing action on the riser tube was connected to both models. It was reasoned that this mixing system would improve gas collection by disturbing the sludge and to improve mixing within the model. Improved mixing was necessary because gas production within the DFF's was not sufficient to prevent the PCS from settling and concentrating at the bottom of the model. As a result, the SRT was significantly lower than the HRT. This created a downward trend in model performance. Unfortunately the mechanical mixer failed to alleviate either of these problems.

Two measures were implemented to resolve the situation. The first was a complete mixing of reactor contents daily, before and after feeding. This was accomplished by raising and lowering the wood media rapidly with a twisting motion which completely mixed the DFF contents (thus increasing the SRT to equal

the HRT), removed any blockage and degasified the sludge. The second measure was to decrease the feed rate from 0.7 l/d to 0.3 l/d yielding increased HRT and SRT which resulted in more noticeable PCS digestion.

The increase in gas production (resulting from improved PCS digestion) was visible, but again not measurable since significant quantities of gas were still being trapped in the sludge. The plexiglas covers were then replaced by plastic funnels in an effort to reduce the headspace above the liquid surface and to provide for media movement to dislodge gas; but gas collection only marginally improved. However, at the beginning of the final month of operation (on day 241) gas production was measured by displacement of model contents (Figure 6.5) and this proved successful. This was accomplished by blocking the biogas collection line; as a result the biogas forced an equal volume of sludge from the model. The displaced sludge was replaced during the next day's feeding. This method was not completely effective for gas collection as it did not recover the gas from the displaced contents.

In an attempt to further improve digestion, 0.6 l of rumen fluid was added to DFF No. 2 (35°C) in the second-last week of operation. The rumen fluid was not added to the 55°C model as the temperature shock would have upset or killed the bacteria in the fluid.

It was hoped that a rumen seed would add many of the microorganisms best suited to degrade cellulosic materials. Gas production volume increased immediately, by 80%, but a large portion of this increase may have been due to the gas production from the rumen fluid itself. The following (last) week's biogas production volume fell, but remained 30% above normal; this suggests that the rumen did in fact provide better seed organisms than were present before.

The fibrous and clumping nature of the PCS also caused sampling and analytical problems. Representative samples of PCS feed and effluent proved difficult to obtain and even more so to analyze, leading to contradictory COD and solids removal results. Consequently, COD destruction was estimated on the basis of methane production.

9.2 Gas Production and COD Removal

COD removal for both models was based on methane production, assuming that 0.35 l of CH_4 were produced for every gram of COD destroyed (theoretical value). Table 9.1 presents the average COD removal and the volume of CH_4 generated per unit of TVS fed to each of the DFF models.

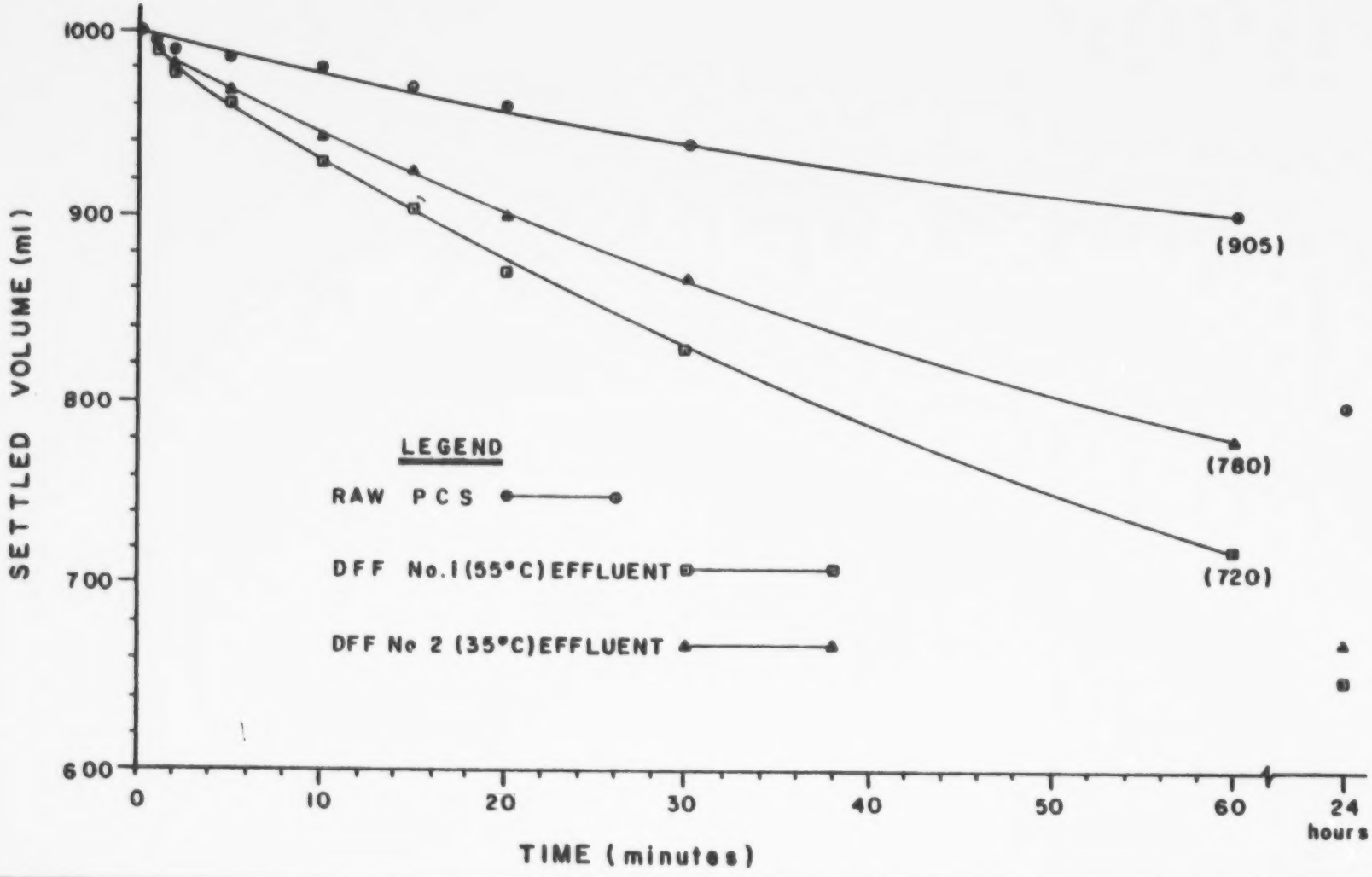
TABLE 9.1 COD REMOVAL AND UNIT CH_4 PRODUCTION (l/g TVS fed)

Model	COD REMOVAL (%)		Unit CH_4 Production (l/g TVS fed)
	Average	Range	
DFF No. 1 (55°C)	24	21-28	0.105
DFF No. 2 (35°C)	26	21-29	0.113

9.3 PCS Settling and Dewatering

Results of two settling tests, in which raw PCS, DFF No. 1 (55°C) effluent and DFF No. 2 (35°C) effluent were allowed to settle in 1 litre graduated cylinders, are presented in Figure 9.1. Settled sludge volume was recorded for one hour and once more after a total of 24 hours.

In addition to the settling tests, the Capillary Suction Time (CST) technique was also used to determine any changes in the dewatering characteristics of the PCS, as a result of anaerobic



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Drawing: PCS SETTLING TESTS			Drawing No.: FIGURE 9.1	

digestion. This technique measures the time required for water (draining from the sludge in question) to move a fixed distance across a filter paper (between two electrical contacts). It has been shown that the rate of movement of water through the filter paper is virtually independent of the sludge depth and is related only to the sludges dewatering characteristics. More specifically the lower the CST, the more easily a sludge may be dewatered. Table 9.2 presents the average CST values for raw PCS, DFF No. 1 (55°C) effluent and DFF No. 2 (35°C) effluent.

TABLE 9.2 SLUDGE DEWATERING RESULTS

Sample	CST (Seconds)
Raw PCS	4.8
DFF No. 1 Effluent	3.8
DFF No. 2 Effluent	4.5

10.0 DISCUSSION OF HYSSL RESULTS

10.1 Model Performance

The BOD removals for the BVF (73.3%), the cloth media DFF (71.0%) and the rock media DFF (67.2%) during the "steady-state" periods indicate that anaerobic treatment of HYSSL is technically feasible. These results are similar to those obtained in previous studies on pulp and paper wastes (1,4,5). The COD removals were lower (29.9-34.2%); however this is typical of pulp and paper wastes since a large portion of the COD material is not easily biodegraded.

Previous studies (4,5) indicate that significant subsequent BOD removal (greater than 70%) is possible from a simple aerated lagoon system. Such a polishing aerobic system would also serve to oxidize the sulphide in the effluent and would reduce sulphide odour and toxicity potential, as well as increase overall BOD removal to 90% amd higher. Experience has shown that aerobic polishing, im the form of biofilters (with high recycle), nitrificatiom basins (activated sludge process) and aerated ponds will not create odour nuisance. The explanation offered for the absence of odour nuisance is the fact that hydrogen sulphide is readily oxidized in the presence of dissolved oxygen and the pH of the aerobic liquor, which follows anaerobic treatment, is usually in the 7.5-8.0 range. In this pH range most of the hydrogen sulphide is in the dissociated form and therefore can not be easily stripped and is therefore oxidized to elemental sulfur and sulfate.

It is also of interest to note that during the last 40 days of operations (Figure 8.6)) when BOD loads were higher than average (1.31 kg BOD/m^3.d for the cloth media DFF) that BOD removals remained high (73.3% - cloth media DFF). This observation

reinforces that of other researchers that treatment performance improves as the anaerobic reactor "matures".

The BOD loads (BVF = 0.42; cloth DFF = 1.00 and rock DFF = 0.81 $kg/m^3.d$) were low compared to those reported for more easily degraded wastes, (e.g., food processing wastes), but are similar to those reported previously for pulp and paper wastes. Normally the average BOD load applied to a high-rate system, such as a DFF, would be five to ten times that applied to a low-rate system, such as the BVF. However, in this study, the DFF BOD loads were only 2.4 (cloth media) and 1.9 (rock media) times that applied to the BVF. One possible explanation is the fact that the BVF was originally seeded with three times as much seed sludge and therefore acclimated faster and was more "mature" than the DFF's. A second and more plausible explanation is that sulphide inhibition/toxicity limited the BOD loading by limiting the HYSSL concentration which could be applied. The average effluent sulphide concentrations (280-350 mg/l, Tables 8.1-8.3) were in the range identified as inhibitory/toxic in previous studies (24,25). Therefore, it is speculated that neither the BVF or DFF's could realize their full loading potential. Certainly the high sodium content of the HYSSL posed no problems since previous research (25) reported satisfactory fermentation at even higher concentrations. The maximum HYSSL concentrations fed to the DFF and BVF models (during steady-state operation) were 40% and 35%, respectively. Earlier in the study when a concentration of 75% was reached, rising sulphide and VA levels made reactor failure imminent in all models. It is felt that higher BOD loads for both the BVF and DFF's could be attained by limiting the HYSSL concentration to, say 20%, and increasing the feed rate accordingly. Alternatively a scrubber-stripper could be utilized to reduce the sulphide levels (see Section 10.3) and recover sulphur.

It is interesting to note that the cloth media DFF appears to have performed somewhat better than the rock filter, having achieved a 23% higher loading (for the same BOD removal) and was more stable at this loading. It can be argued that this small difference in performance is really not significant. However, it can be explained by a number of important factors, or their combinations, viz:

1. the SSA of the cloth was approximately 30% higher, i.e., 115 vs. 90 m^2/m^3;

2. the gas scrubber-stripper was installed on the cloth filter and not the rock - this lowered the sulphide concentration and stress in the cloth media DFF;

3. the cloth media, because of its wooly surface, has superior seed sludge retention and thus provides superior organism-substrate contact;

4. the cloth DFF had over 200% more HRT due to the fact the rock displaces almost 60% of the liquid volume; and

5. the rock filter accumulated more solids than the cloth model (as evidenced by SS results) which would make the relative HRT's worse still, although the accumulated solids contain bacteria which, of course, degrade the waste.

Previously, it was stated that sulphide inhibition/toxicity may have been causing system stress in all reactors, to the extent that it seriously limited their loadings. This is also evidenced by the fact the effluent VA levels were considerably higher than those normally seen in a well operating, mature anaerobic system. However, previous studies (26,27) have revealed that stable operation at high VA levels is possible.

At the completion of the laboratory study, the DFF's were opened for inspection and several important observations were made:

1. the very top section of the cloth DFF had a thick buildup of sludge which may have prevented access of influent to some portions of the filter;

2. with the exception of the very top section, the film on the cloth media was quite uniform in thickness (approximately 2-3 mm on both sides) and appearance from top to bottom; and

3. the top section of the rock filter contained a thick accumulation of sludge in all the interstitial spaces which then appeared to thin out towards the bottom of the filter.

Of the three models, only the BVF effected significant SS removal (81.9% versus 7.5% and a negative removal for the rock and cloth DFF's, respectively). The long HRT and horizontal flow mode of the BVF encourages solids settling and subsequent digestion of volatile solids.

In every anaerobic filter design, the means to handle the accumulation of suspended solids from the influent feed or from the generation of biomass, must be provided. This is true even in the case of a totally soluble influent as biomass will be generated and may accumulate to the point of affecting system performance. If effluent solids create a problem for subsequent processes or with the discharge permit, a subsequent solids separation device will be required. The downward flow mode of the DFF encourages the wash out of solids with the effluent. It was, in fact, this phenomenon which prompted the development of this technology for it has the advantage of reducing the possibility and problems associated with media clogging, but the

disadvantage of increasing effluent solids (with attendant increase in COD and BOD). The rock media DFF SS removal was only 7.5%, but for the cloth media DFF average effluent SS was 1 280 mg/l greater than the influent. This was due to two factors; the straight vertical channel pores of the cloth media (Figure 6.3) did not "trap" many of the influent solids and the fact that media biofilm will only develop to a limited thickness before sloughing bios. A solids separation device would be required to remove suspended solids from the cloth DFF effluent prior to aerobic treatment or discharge. Assuming such a device would yield SS values similar to those produced by BVF, the clarified DFF effluent BOD would have been appreciably lower and the removal higher. Furthermore, provisions for sludge recycle should improve allowable loadings and/or performance as well.

The rock media, on the other hand, had much smaller pore spaces and no straight vertical channels through the media; media of this type would have a tendency to trap solids and also allow greater accumulation of biological solids. This suggests that careful attention has to be paid to the design of the solids removal system, not only for a rock filter, but any type of media.

The specific methane production rates for all three models (BVF = 0.52, cloth media DFF = 0.56 and rock media DFF = 0.47 m^3 CH_4/kg BOD removed) were similar to results in a previous study (5). The specific CH_4 rates expressed on a COD removed basis ranged from 0.21 to 0.26 m^3 CH_4/kg. These values are well below the theoretical value of 0.35 m^3 CH_4/kg COD removed. Two reasons for this are the reduction of various oxidized forms of sulphur to sulphide (28) and the conversion of COD to biomass (which was still accumulating in these immature systems). The high feed sulphur content also resulted in high H_2S gas concentrations (BVF = 8.5%, cloth media DFF = 3.9% and

rock media DFF = 8.8%) in the biogas. The gas scrubber connected to the cloth media DFF on day 219, resulted in a considerable drop in its biogas H_2S (and CO_2) content (from 8 to about 1.5% H_2S) and a corresponding increase in CH_4 content (from 56% to about 90%). It should also be pointed out that not all the biogas passed through the scrubber as a gas storage bag preceded it; if all the biogas had been scrubbed the resultant composition of the biogas would have indicated almost pure methane. Therefore to control sulphide in the anaerobic effluent and biogas the HYSSL should be diluted with a low sulphur waste stream prior to treatment (water was used for dilution during this study) or the gas scrubbed and recycled to provide sulphide stripping (as in the cloth media) DFF, (see Sections 8.5 and 10.3 which discuss sulphur balance).

10.2 Dynamic Testing

The results of the cloth model DFF feed cessation (and subsequent resumption) are illustrated in Figure 8.17. Before feed interruption the steady-state biogas production rate was 5.0 l/d. After 12 hours of no feed the gas production rate fell to 0.7 l/d, and decreased further to 0.5 and 0.4 l/d after 36 and 60 hours, respectively. Only 18 hours after feed resumption the gas production rate returned to its former steady-state rate.

The results of the shock load program during which the flow to the cloth media DFF was tripled (from 1.2 to 3.6 l/d) are presented in Figures 8.18 and 8.19. This triple load was applied on day 274. During the week before this experiment the model had been under some stress, as evidenced by high VA (approximately 2100 mg/l) levels (Figure 8.10). On day 274 the VA had fallen to 1300 mg/l, but as the shock load was applied it rose rapidly (to 2 650 mg/l in 24 hours) and continued upwards even after the flow returned to normal (1.2 l/d). By day 279 the VA

had increased to 4 050 mg/l and the model was shut down as the study period had expired. The fact that the model showed no sign of recovery following the shock load indicates that it was operating under stress and near its full capacity before the shock loading was applied.

The decrease in CH_4 Production, shown in Figure 8.18, also reveals the DFF's stressed condition following the shock load. Immediately after the shock loading CH_4 Production fell from 3.3 l/d to 1.4 l/d. As the VA levels continued upwards CH_4 production fell further so that three days later it was 0.7 l/d. The CH_4 Content of the biogas was also adversely affected by the load increase.

To conduct this experiment the gas scrubber was removed so that increases in CO_2 Content of the gas could be detected. During the shock loading the biogas CH_4 Content was 46% and this decreased to 22% and 11% 48 hours and 96 hours later, respectively. Decreases in CH_4 Content (and corresponding increases in CO_2) are typical in highly-stressed anaerobic reactors. To recover from such a stress the model would have to be shut down to allow the VA levels to return to normal or perhaps the alkalinity increased to ensure proper pH levels while the the excess VA is consumed in the reactor. The sulphide toxicity factor would also have to be taken into account.

10.3 Sulphur Balance

The results presented in Section 8.5 reveal that over a three week test period using HYSSL at 35% concentration, 35.1 g of sulphur was fed into the BVF. Over the same period a total of 32.0 g left (19.7 g in the treated effluent and 12.3 g in the biogas). Therefore, approximately 91% of the sulphur fed was accounted for. The remaining 9% may be accounted for in sedimentation of feed solids in the BVF, sulphur required for biological growth (cell production), lag time (influent and

effluent samples were collected over identical periods and the BVF HRT was 11.1 days) and experimental error. In any case, this data indicates a very significant portion of the sulphur in the spent liquor was transformed into hydrogen sulphide and expelled from the reactor via the biogas stream.

In similar tests with the cloth media DFF, (using HYSSL at 40% concentration) the removal of sulphur from the spent liquor was even greater. In this case, the results of a two week testing period revealed that while 28.0 g of sulphur was fed to the DFF, only 9.4 g left in the liquid effluent stream. It is believed that the high level of sulphur removal (66%) resulted from the installation of a crude biogas scrubber-stripper. The scrubber-stripper was used to produce a relatively pure stream of CH_4 gas by removing the CO_2 and H_2S. This was done by bubbling the biogas through caustic solution. The methane rich gas was then recycled through the reactor to strip more of the sulphide for subsequent removal in the scrubber. The purpose of this system was to lower sulphide levels in the reactor and thereby reduce the potential for sulphide toxicity/inhibition.

Further analysis of the results suggest that there is potential for even higher levels of sulphur recovery from the effluent as a large portion of the total sulphur was in the soluble sulphide form. In a properly designed sulphide stripping system using proper pH control, the sulphide stripping efficiency should be improved from that achieved here with this very crude system.

ADI Limited will begin, very shortly, an anaerobic treatability study (for the Environmental Protection Service) using Kraft foul condensates. This study will assess the effectiveness of a biogas scrubber-stripper equipped DFF for the removal of BOD and Total Reduced Sulphur (TRS) from the foul condensates. This

novel system may prove to be a cost-effective way of accomplishing TRS and BOD removal in the same system.

Another possible benefit of the biogas scrubber-stripper concept is the potential for sulphur recovery. By using sodium hydroxide solution as the scrubbing liquid, the spent liquid will contain a mixture of NaHS, Na_2S, Na_2CO_3 and excess NaOH. In this form there are chemical recovery processes available which will permit recycle of chemical into the pulp cooking liquor (both Kraft and sulfite). More work is certainly required in the whole area of anaerobic treatment of spent sulphite liquors with sulphur recovery before the preliminary economics can be refined. However, as the price of sulphur rises (as it has been over the past few years), so does the potential for economical recovery using this novel process.

In addition to the potential financial benefits of sulphur recovery, there are those on the environmental side, viz:

(1) less sulphur in pulp and paper wastes translates into less potential for odour nuisance in receiving streams, and

(2) less potential for toxicity toward fish (in the receiving stream).

11.0 DISCUSSION OF PCS RESULTS

11.1 Gas Production and COD Removal

For reasons detailed in Section 9.1 COD removals were estimated based on CH_4 production. The average COD removals, based on CH_4 Production, for DFF No. 1 (55°C) and DFF No. 2 (35°C) were 24% and 26%, respectively. These results are based on the period from day 241 to 276, since before day 241 difficulties in gas collection prevented accurate measurement of biogas (Section 9.1). These results suggest that, in terms of sludge digestion (COD/TVS destruction), there are no obvious benefits in operating in the thermophilic range.

The difficulties encountered in operation of the DFF models suggest that for a full-scale PCS digestion system and similar applications to sludges where solids separate during digestion, the DFF would probably not be the most appropriate technology. The nature of the PCS (especially after it begins digesting) was such that the DFF could not provide adequate mixing (through gas production) to keep the PCS from settling to the bottom of the reactor. This phenomena resulted in very short SRT and thus poor PCS digestion. A completely mixed process, such as a contact digester, would probably be most appropriate. This type of reactor would be able to maintain a high inventory of anaerobic organisms through sludge recycle and promote good mixing and contact between the organisms and the PCS. The economics of full-scale PCS digestion, in Section 12.2, were based on a contact digester achieving 30% COD removal. For this case the contact digester appears to be very cost-effective. A capital cost was developed for a DFF. However, it was found to be considerably more expensive than the contact digester and dropped from further consideration.

The slight increase in COD removal with the contact digester (over the DFF models) is justified on the basis of better mixing (completely mixed), sludge recycle, better innoculum and greater quantity of anaerobic organisms. In this regard, it should also be mentioned that the full digestion potential of these DFF models was never realized. This was due to the fact that the media was used as a mixing element and a good growth of biomass never developed. At best the biological film on the wood media could only be described as scanty.

The DFF would be applicable for digestion of most organic sludges, since most would not have the fibrous, clogging characteristics of the PCS. As well, PCS digestion could be more easily accomplished if it were mixed with a more easily digested material (29) (i.e. secondary sludge). It is very likely that anaerobic digestion will be effective on most types of organic sludges common to the pulp and paper industry and thereby may offer the industry a new and cost-effective approach to sludge disposal.

11.2 PCS Settling and Dewatering

The results of the PCS settling tests, for raw and digested samples presented in Figure 9.1, indicate 14% and 20% reductions in settled PCS volume by DFF No. 2 (35°C) and DFF No. 1 (55°C), respectively. These percentages are based on results after 60 minutes of settling. Such waste volume reductions are important to the waste treatment operator in terms of reduced disposal cost (particularly if the waste is not dewatered prior to disposal). Improvements in settling and dewatering characteristics are typical for anaerobically digested sludges.

Results of the CST dewatering tests also indicate improvements in PCS dewatering characteristics. The CST times were reduced 6% and 21%, in DFF No. 2 and No. 1, respectively. The dewatering improvements are also of importance to the full-scale operator in terms of reduced operational costs, drier (better burning) cake and reduced volume for landfill.

In terms of improvement in sludge dewatering, thermophilic digestion resulted in some advantages over mesophilic operation. However, these improvements may be more than off-set by two disadvantages. The first is the higher energy input that may be necessary to sustain the 55°C digester temperature and secondly, thermophilic digesters are typically more difficult to operate and control as the bacterial population (at 55°C) is more sensitive to environmental changes (pH, temperature, etc.). When all the advantages and disadvantages are weighed, mesophilic operation is preferred.

12. PROJECTED FULL-SCALE ECONOMICS

12.1 High Yield Spent Sulphite Liquor

The economics for the HYSSL treatment were developed around a hypothetical sodium base, sulphite mill with a yield of approximately 65% and production of 250 tonnes (oven dried) of pulp per day. It is assumed that the spent liquor contains approximately 50 000 kg BOD/d (i.e., approximately 200 kg BOD/oven-dried tonne) and is discharged without treatment into a local stream.

The capital cost for an anaerobic system to treat the HYSSL was developed for only the ADI-BVF technology. This was done for several reasons, viz.:

(1) the BOD/COD removal in the BVF was marginally better than those in the DFF's; and

(2) experience has shown that in order to be cost effective, the DFF BOD load must be much higher than in the BVF; in this study the ratio was less than 3 to 1.

The estimated capital cost for an ADI-BVF, to remove approximately 75% of the HYSSL BOD (i.e., approximately 37 500 kg/day) is $4 150 000. This cost is for a completely installed system including allowances for interfacing with the pulp mill and an outfall. The complete economic summary is presented below.

HYSSL Economic Summary

		$ 1983
1.	Capital Cost (ADI-BVF)	$4 150 000
2.	Annual Operating and Maintenance Costs	271 000
3.	Annual Biogas Revenue	1 400 000
4.	Net Savings (3-2)	1 129 000

Simple Payback (1 divided by 4) = 3.7 years

From the above table, it can be seen that the simple payback for the BVF system is attractive. This is especially true for the mill that is now providing external biological (aerobic) treatment or the mill which is being forced into it. In these cases, the operating and maintenance costs for the existing or alternative aerobic treatment system, appear as savings attributable to the anaerobic system. These additional savings (which will be quite large as they will include a significant amount for aeration power) will naturally produce an even better payback - 2 to 3 years being typical.

In any case, this analysis points to the fact that anaerobic treatment of spent sulphite liquors may offer an attractive return on investment, while solving a long-standing pollution problem for many sulphite mills in Canada.

12.2 Primary Clarifier Sludge

The projected costs, revenues and savings for an anaerobic system to treat primary clarifier sludge were based on a hypothetical mill producing approximately 800 oven dried tonnes of newsprint per day. It was assumed that this mill had total suspended solids losses of approximately 36 tonnes per day of

which 32 tonnes per day are recovered as primary sludge at approximately 1% solids.

A number of system scenarios exist for disposal of primary sludge, four of which are:

(1) vacuum filtering followed by landfilling;
(2) vacuum filtering followed by pressing and incineration;
(3) anaerobic treatment followed by number 1 above (with smaller vacuum filtering and landfill requirements); and
(4) anaerobic followed by number 2 above (with smaller dewatering facilities and incineration equipment).

For purposes of this report, it was assumed that the hypothetical mill is presently vacuum filtering and landfilling its filter cake (scenario 1) and assessing the cost-effectiveness of modifying its system to scenario 3. Since the mill was assumed to exist, there was no capital cost credit for a smaller vacuum filtering system and any savings in operating and maintenance costs was ignored.

There were many operating difficulties encountered with the bench-scale DFF's while treating PCS. Many of these problems may not have occurred if a contact digester had been used, which for this particular application would have been much more suitable. It was only during the last month of testing that meaningful data were collected and it was on this limited data that capital cost estimates were prepared.

The basic design criteria established from the last several weeks of PCS testing were:

(1) there appears to be little benefit to operate in the thermophilic (55°C) range and therefore 35°C is to be used;
(2) COD/TVS loading = 0.87/0.68 kg per m^3.d (based on liquid volume and not liquid plus media); and
(3) COD and TVS destroyed, at least 30%.

The above TVS loading is believed to be conservative, especially for some sludges and sludge mixtures, as it is in the lower range for standard rate municipal type anaerobic digesters. Work reported by Takeshita et al (29) at 55°C on pulp and paper wastes also indicates the above design criteria are conservative, even after allowance for higher temperature. These researchers achieved volatile suspended solids (approximately equal to total volatile solids in the case of sludges) destruction and loadings of the order of 50 to 70% and 3 to 4 kg/m^3.d, respectively, using the anaerobic contact process.

Capital costs were developed for both the DFF and contact digester using the design criteria stated in No. 2 and 3 above. The capital costs for the DFF and contact digester (C.D.) are $11.0 and $3.6 million, respectively (reference Appendix 3). The major reasons the contact digester is significantly lower in cost are:

(1) the C.D. has no media (media is very expensive); and
(2) the feed and outlet header systems are less complex in the C.D.

A summary of the capital and operating and maintenance costs, biogas revenue and landfilling savings is presented below for the C.D. only.

PCS Economic Summary

		$ 1983
1.	Capital Cost (contact digester)	3 625 000
2.	Operating and maintenance costs	239 000
3.	Biogas revenue	287 000
4.	Landfill savings	896 000
5.	Net Savings (3 + 4 - 2)	$ 944 000

Simple Payback (1 divided by 5) = 3.8 years.

A complete list of all data and assumptions, and detailed cost revenue and savings calculations appear in Appendix 3.

The simple payback for the PCS contact digester system is approximately 4 years, which is quite attractive. It would therefore appear that the anaerobic sludge digestion route for many pulp and paper sludges (not only PCS from a newsprint mill) may offer a new and cost-effective approach to sludge disposal. However, it should be pointed out that the economics are quite sensitive to such factors as degree of solids digestion and improvement in cake solids and hauling and landfill costs.

For mills that do not practice landfilling as a means of ultimate sludge disposal, anaerobic digestion to reduce sludge volumes can still be advantageous. The reduction in sludge volume comes as a direct result of digestion and also through an improvement in sludge dewatering characteristics. This reduction in sludge volume and sludge water content could produce the following benefits:

(1) smaller, less costly dewatering facilities or less operating hours on existing facilities;
(2) drier cake for incineration;
(3) higher capacity from wood waste boiler and improved thermal efficiency of same.

For the above analysis it was assumed that the improvement in the sludge dewatering characteristics would increase the vacuum filter cake solids from 20 to 30%. There is little justification from the results presented here or from data in the literature, to substantiate this increase in cake solids. However, it is believed to be realistic and certainly achievable with chemicals purchased from the savings in operating and maintenance costs because of lower sludge volume. It was also stated previously that the COD/TVS destruction was assumed to be at least 30% and that other work (29) on pulp and paper sludge suggests that higher loadings and volatile suspended solids destruction of 50 to 70% is reasonable to expect. This wide range of solids destruction (biodegradability) is a function of the sludge composition, the more highly-degraded sludges having a lower lignin content and vice-versa.

It should also be pointed out that where waste secondary sludge is available, it would be advantageous to anaerobically digest it along with primary sludges as it is more biodegradable and will reduce the need for nutrients.

13.0 CONCLUSIONS AND RECOMMENDATIONS

13.1 Conclusions

1. The following table summarizes the BOD and COD loads and removals obtained in the ADI-BVF, cloth media DFF and rock media DFF reactors treating HYSSL.

HYSSL Data Summary

Reactor	BOD		COD	
	Load ($kg/m^3.d$)	Removal (%)	Load ($kg/m^3.d$)	Removal (%)
ADI-BVF	0.42	73.3	2.48	34.2
Cloth DFF	1.00	71.0	5.73	29.9
Rock DFF	0.81	67.2	4.61	33.0

2. Of the three HYSSL models only the BVF achieved significant SS removal (81.9%, compared to 7.5% and no removal for the rock and cloth media DFF's, respectively). The horizontal flow mode and long HRT of the BVF encourages solids settling, whereas the DFF has a tendency to expel solids with the effluent.

3. All of the HYSSL models acclimated to relatively high levels of sulphide (with influent values as high as 1 000 mg/l).

4. The following table presents the average specific CH_4 production rates and biogas composition for the HYSSL models. The observed rates are less than the theoretical value due to reduction of various forms of sulphur to sulphide and production of biomass (and possibly other factors).

HYSSL Model Biogas Data

Reactor	Specific CH_4 Rate m^3/kg BOD (COD) removed	Biogas Composition (%) CH_4	CO_2	H_2S
ADI-BVF	0.52 (0.25)	51	40.5	8.5
Cloth DFF	0.56 (0.26)	69*	27.1	3.9
Rock DFF	0.47 (0.21)	49	42.2	8.8

*The cloth media DFF employed a biogas scrubber/stripper during the last 53 days of operation which decreased the average H_2S and CO_2 concentrations (and increased CH_4).

5. The cloth media DFF exhibited rapid response (in terms of biogas production) to cessation and subsequent resumption of feed.

6. Methane production decreased rapidly (and VA concentration increased accordingly) following a 24 hour shock load (flow tripled) to the cloth media DFF. The model showed no sign of recovery following the shock input indicating it was already under stress and operating at near full capacity.

7. BOD loadings to all HYSSL models appeared to be limited by sulphide inhibition/toxicity at HYSSL feed concentrations of 35-40% full strength. Lower concentration may have resulted in higher loadings. While sodium concentrations were high they were judged not to be a problem on the basis of previous research. However, the synergistic effects of sodium, sulphide and other ions, as they relate to toxicity/inhibition, are unknown.

8. A sulphur balance conducted for the BVF indicated that 91% of the sulphur fed was accounted for in the effluent and biogas. A similar analysis for the cloth media DFF revealed that only 34% of the sulphur input left in the effluent and biogas. Most of the remaining (66%) sulphur was removed by a crude biogas scrubbing/stripping system. With a properly designed sulphide stripping system even higher levels of sulphur could have been removed from the effluent which open up interesting possibilities for chemical recovery.

9. Operation of the two wood media PCS DFF's proved difficult, owing to the nature of the PCS. Problems included clogging, within model settling (resulting in solids loss, i.e., low SRT) poor mixing and difficulties in collecting and measuring biogas.

10. Based on the quantities of CH_4 produced the PCS COD removals from DFF No. 1 (55°C) and DFF No. 2 (35°C) were 24 and 26%, respectively.

11. Anaerobic fermentation resulted in improved PCS settling and dewatering characteristics.

12. On the basis of the results of this study it is felt that anaerobic fermentation should find wide application in the pulp and paper industry in the future.

13. In spite of the apparently low loadings achieved during steady-state operation, the projected full-scale economics for anaerobic fermentation of HYSSL and PCS appear promising. The tables below summarize full-scale costs and savings for anaerobic fermentation of HYSSL and PCS (details Section 12.0).

HYSSL Economics*

(1) Capital cost - ADI-BVF	$4 150 000
(2) Annual operating and maintenance cost	271 000
(3) Annual biogas revenue	1 400 000
(4) Net annual savings	1 129 000

Simple payback = (1)/(4) = 3.7 years

*Hypothetical case - sulphite mill with 65% yield producing 250 tonnes of pulp per day.

PCS Economics*

(1) Capital cost - contact digester	$3 625 000
(2) Annual operating and maintenance costs	239 000
(3) Biogas revenue	287 000
(4) Landfill savings	896 000
(5) Net annual savings	944 000

Simple payback = (1)/(5) = 3.8 years

*Hypothetical case - newsprint mill, production = 800 tonnes per day.

13.2 Recommendations

The following list recommends topics for further work relevant to application of anaerobic fermentation to pulp and paper wastes.

1. Pilot-scale work is necssary with HYSSL to:

   (a) explore increased loads with more dilute HYSSL and develop full-scale design criteria;

(b) simulate day-to-day fluctuations of waste characteristics and flow and effects on performance;

(c) determine the full potential for sulphur recovery through sulphide scrubbing and stripping with the use of a proper scrubber-stripper system;

(d) investigate alternative systems for chemical recovery from scrubber liquor and the potential benefits and economics;

(e) determine nutrient and alkalinity requirements;

(f) assess the impact of aerobic polishing regarding sulphide and further BOD removal; and

(g) assess the toxicity of the aerobic effluent.

2. A comprehensive survey of all pulp and paper waste streams and sludges should be undertaken to determine the potential for energy conservation through anaerobic treatment and other possible benefits.

3. Bench-scale (followed by pilot) work is needed to better quantify the TVS loadings and destruction, and the effects on dewatering characteristics by anaerobic digestion of PCS and other pulp and paper sludges.

4. The potential means/problems of biogas utilization with high sulphide concentration should be investigated and means of upgrading where necessary.

5. A study should be undertaken with a spent liquor of low sulphur content to determine if sulphur forms are the primary cause for toxicity/inhibition and low organic loading rates (ADI is already in the planning stages for this work).

APPENDIX I

GROSS ENERGY SAVINGS PER TONNE COD DIGESTED ANAEROBICALLY

APPENDIX 1

Gross Energy Savings Per Tonne COD Destroyed Anaerobically

Data and Assumptions:

1. 0.35 m^3CH_4 produced per kg COD destroyed
2. 37 000 kJ/m^3CH_4
3. BOD/COD ratio = 0.35 (typical for pulp and paper wastes)
4. 1.2 kg O_2 required per kg BOD destroyed
5. 1.0 kg O_2/kWh oxygen transfer rate
6. Electrical energy production - 35% efficient (fossil fuel)
7. $5.50/GJ

Energy content of biogas:

$$1000 \text{ kg COD} \times \frac{0.35 \text{ m}^3\text{CH}_4}{\text{kg COD}} \times \frac{37\,000 \text{ kJ}}{\text{m}^3\text{CH}_4} = 13.0 \text{ GJ}$$

Electrical Energy Savings

$$1000 \text{ kg COD} \times \frac{0.35 \text{ kg BOD}}{\text{kg COD}} \times \frac{1.2 \text{ kg } O_2}{\text{kg BOD}} \times \frac{\text{kWh}}{1.0 \text{ kg } O_2} \times \frac{1}{.35} \times \frac{3.6 \text{ MJ}}{\text{kWh}}$$

$$= 4.3 \text{ GJ}$$

Gross value of biogas and electrical energy savings per tonne COD destroyed:

$$(13 + 4.3) \text{ GJ} \times \$5.50/\text{GJ} = \$95.00/\text{tonne COD}$$

APPENDIX 2

HIGH YIELD SPENT SULPHITE LIQUOR

APPENDIX 2

HIGH YIELD SPENT SULPHITE LIQUOR

1. DESIGN CRITERIA AND ASSUMPTIONS

a.* BOD/SS - 12 400/910 mg/l
b.* Quantity of liquor - 4 000 m^3/d
c.* TKN - 48 mg/l
d.* Total phosphorous - 5 mg/l
e.* Temperature - 90 to 100°C
f. Dilution required - 1 part HYSSL/2 parts washwater
g. Temperature after dilution - 40°C
h. Reactor type - ADI-BVF
i. BVF loading/(volume) - 0.5 kg BOD/m^3.d/(100 000 m^3)
j. BOD removal - 75%
k. Methane yield - 0.52 m^3/kg BOD destroyed
l. BOD/N/P required - 500/5/1
m. Chemicals for alkalinity control - nil
n. Mill operation - 350 d/a
o. Value of Biogas - $5.50/GJ
p. Cost of electricity - $0.04/kWh
q. land is available with no unusual site conditions
r. existing boilers converted to burn biogas with high sulphur content
s. all biogas produced can be burned (i.e., no storage required)

* before dilution

2. ESTIMATED CAPITAL COST FOR HYSSL - ADI-BVF

		$ 1983
a.	site preparation, excavation, backfill, clay liner and subdrains	850 000
b.	yard piping, inlet and outlet headers, sludge recycle and waste system, force main and gravity outfall	425 000
c.	insulated, floating membrane cover	1 300 000
d.	biogas system including electrical, duplex blowers, blower house, controls, gas flare, emergency vent, instrumentation, gas main to plant, and boiler modifications	500 000
e.	wastewater pumping station, allow	100 000
f.	nutrient addition, allow	100 000
g.	monitoring and controls, allow	30 000
h.	seed sludge, allow	20 000
	Subtotal	$3 325 000
i.	contingencies and engineering @ 25%	825 000
	Total	$4 150 000

3. ESTIMATED ANNUAL OPERATING AND MAINTENANCE COSTS

a.	operator-technician - 1 shift position		25 000
b.	electrical costs (@ $0.04/kWh)		9 000
	raw waste pumps, allow	10 kW	
	recycle pumps, allow	8 kW	
	gas blower, allow	8 kW	
	Total	26 kW	
c.	chemicals		
	alkalinity		Nil
	nitrogen 240 t/a (as urea) @ $490/t		118 000
	phosphorous 146 t/a (as super-P) @ $245/t		36 000
d.	maintenance @ 2% of total capital cost		
	0.02 x $4.15 million		83 000
	Total		$ 271 000

4. BIOGAS REVENUE

$$\frac{4\ 000\ 000\ l}{d} \times \frac{12\ 400\ mg\ BOD}{l} \times \frac{kg}{10^6\ mg} \times 0.75$$

$$\times \frac{0.52\ m^3\ CH_4}{kg\ BOD} \times \frac{0.037\ GJ}{m^3 CH_4} \times \frac{\$5.50}{GJ} \times \frac{350\ d}{a}$$

$$= \$1\ 400\ 000/a$$

APPENDIX 3

PRIMARY CLARIFIER SLUDGE

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PRIMARY CLARIFIER SLUDGE

1. DESIGN CRITERIA AND ASSUMPTIONS

a. COD/TS/TVS = 13 000/10 600/10 200, mg/l
b. TKN = 14 mg/l
c. Total phosphorous = 1 mg/l
d. Primary sludge - 32 tpd (total dry solids)
e. Total solids/total volatile solids = 1.0%/0.96%
f. Temperature of sludge leaving clarifier - 40°C
g. Reactor temperature - 35°C
h. Reactor loadings - 0.87 kg COD/m^3.d or 0.68 kg TVS/m^3.d (these loadings are based on liquid volume, not liquid plus media)
i. COD and TVS destroyed - approx. 30%
j. COD/BOD ratio - approx. 3.5
k. BOD/N/P ratio required - 500/5/1
l. Methane yield - 0.35 m^3CH_4/kg COD destroyed
m. Chemicals for alkalinity control - Nil
n. Mill operation - 350 days/annum
o. Value of biogas - $5.50/GJ
p. Cost of electricity - $0.04/kWh
q. Vacuum filter cake - 20% TS before digestion and 30% after digestion
r. Annualized hauling and landfilling cost for dewatered PCS - $30.00/tonne (30)
s. PCS haul distance - 8 kilometers (30)
t. PCS disposal method - wide trench (30)
u. Exchange on US $ - $1.20 Canadian
v. Inflation rate - 10%/a

2a. ESTIMATED CAPITAL COST FOR PCS CONTACT DIGESTER

		$ 1983
a.	Site preparation, excavation, backfill, clay liner and subdrains	450 000
b.	Yard piping, inlet and outlet headers, force main and gravity sewer, manholes, etc.	200 000
c.	Insulated floating membrane cover	650 000
d.	Inclined plate clarifier and sludge recycle	300 000
e.	Mixers and electrical	600 000
f.	Biogas system including electrical duplex blowers, blower house, controls, gas flare, emergency vent, instrumentation, gas main and boiler modifications	450 000
g.	Wastewater pumping station, allow	100 000
h.	Nutrient addition, allow	100 000
i.	Monitoring and controls, allow	30 000
j.	Seed sludge, allow	20 000
	Subtotal	2 900 000
k.	Contingencies and engineering add 25%	725 000
	Total	$3 625 000

2b. ESTIMATED CAPITAL COST FOR PCS, DFF

		$ 1983
a.	Site preparation, excavation, backfill	175 000
b.	Concrete footings, floor and walls	1 990 000
c.	Media and supports	4 900 000
d.	Yard piping, inlet and outlet headers	700 000
e.	Floating membrane cover	335 000
f.	Biogas system including, electrical, duplex blowers, blower house, controls, gas flare, emergency vent, instrumentation, gas main and boiler modifications	450 000
g.	Wastewater pumping station, allow	100 000
h.	Nutrient addition, allow	100 000
i.	Monitoring and controls, allow	30 000
j.	Seed sludge, allow	20 000
	Subtotal	8 800 000
k.	Contingencies and engineering add 25%	2 200 000
	Total	$11 000 000

3. ESTIMATED ANNUAL OPERATING AND MAINTENANCE COSTS*

			$ 1983
a.	Operator - technician - 1 shift position		25 000
b.	Electrical costs (at $0.04/kWh)		111 000
	sludge recycle pump, allow	8 kW	
	influent pumps, allow	15 kW	
	gas blower, allow	8 kW	
	mixers, allow	300 kW	
	Total	331 kW	
c.	Chemicals		
	alkalinity		Nil
	nitrogen 50 t/a (as urea) @ $490/t		24 000
	phosphorous 28 t/a (as super-P) @ $245/t		7 000
d.	Maintenance @ 2% of total capital cost		
	0.02 x $3.62 million		72 000
	Total		$239 000

*for Contact Digester

4. VALUE OF BIOGAS

$$\frac{32\,000 \text{ kg TS}}{\text{d}} \times \frac{1.2 \text{ kg COD}}{\text{kg TS}} \times 0.3 \times \frac{0.35 \text{ m}^3\text{CH}_4}{\text{kg COD}}$$

$$\times \frac{0.037 \text{ GJ}}{\text{m}^3\text{CH}_4} \times \frac{\$5.50}{\text{GJ}} \times \frac{350\text{d}}{\text{a}} = \$287\,000/\text{a}$$

5. SAVINGS IN LANDFILL COSTS

Present landfill costs:

$$\frac{32\text{t}}{\text{d}} \times \frac{1}{0.20} \times \frac{350\text{d}}{\text{a}} \times \frac{\$30}{\text{t}} = \$1\,680\,000/\text{a}$$

Landfill costs after anaerobic digestion:

$$\frac{32\text{t}}{\text{d}} \times .70 \times \frac{1}{0.30} \times \frac{350\text{d}}{\text{a}} \times \frac{\$30}{\text{t}} = \$784\,000/\text{a}$$

Landfill Savings:

$1 680 000 - $784 000 = $896 000/a

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